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by

Alan F. Cassel
and
Berinda P. Johnson

District of Columbia Government
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Water Resources Management Administration
Washington, D.C. 20032

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Project Officer

Roland V. Villiers
Wastewater Research Division
Municipal Environmental Research Laboratory
Cincinnati, Ohio 45268

MUNICIPAL ENVIRONMENTAL RESEARCH LABORATORY
OFFICE OF RESEARCH AND DEVELOPMENT
U.S. ENVIRONMENTAL PROTECTION AGENCY
CINCINNATI, OHIO 45268

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FOREWORD

The Environmental Protection Agency was created because of increasing public and government concern about the dangers of pollution to the health and welfare of the American people. Noxious air, foul water, and spoiled land are tragic testimony to the deterioration of our natural environment. The complexity of that environment and the interplay between its components require a concentrated and integrated attack on the problem.

Research and development is that necessary first step in problem solution and it involves defining the problem, measuring its impact, and searching for solutions. The Municipal Environmental Research Laboratory develops new and improved technology and systems for the prevention, treatment, and management of wastewater and solid and hazardous waste pollutant discharges from municipal and community sources, for the preservation and treatment of public drinking water supplies, and to minimize the adverse economic, social, health, and aesthetic effects of pollution. This publication is one of the products of that research; a most vital communications link between the researcher and the user community.

This report presents the results of a year and a half study of the capability of various mechanical dewatering devices to produce high solids sludge cake. Results show that a number of alternative methods are presently available that are capable of dewatering municipal sludge solids to 30-40 percent cake dryness. This is quite significant since it impacts in subsequent cost savings in disposal of sludge solids. At 30 percent solids, sludge burns autogenously. This eliminates the need of costly auxiliary fuel to incinerate sludge. Also, high cake solids means less sludge to haul and land dispose. This decreases land disposal costs.

Francis T. Mayo
Director
Municipal Environmental Research
Laboratory, Cincinnati

ABSTRACT

Pilot-scale dewatering tests were made to establish design and operating parameters for dewatering municipal wastewater sludges on recessed plate filter presses (both diaphragm and fixed volume types), continuous belt presses, and retrofit units for a vacuum filter. Results from the 1.5-year study showed that when dewatering lime and ferric chloride-conditioned sludges, the recessed plate presses consistently produced a 30-40% solids filter cake. Feed solids to the units averaged 5% total solids with a range from 2.4 to 10%. Various ratios of waste-activated to primary sludge solids, with emphasis on the 2/1 ratio, were tested. Successful operation of the belt presses on the Blue Plains sludge was largely a function of the percentage of waste activated sludge in the feed mixture. Cake solids from 25-30% were attained when the polymer conditioning dosage was optimized. When used as a retrofit device to a vacuum filter, the belt press gave cake solids in the 30-40% range during laboratory-scale tests. Full-scale demonstration, however, was not achieved because an adequate system for delivering filter cake to a belt filter has not yet been developed.

Design parameters are developed to dewater a mixture of 67% secondary and 33% primary sludge in a full-scale plant installation. The estimated costs for dewatering plus final disposal by either incineration or composting are also presented.

This report was submitted in fulfillment of Contract No. 68-03-2455 by the Water Resources Management Administration, Department of Environmental Services, District of Columbia, under the sponsorship of the U. S. Environmental Protection Agency.

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ABBREVIATIONS AND SYMBOLS

ABBREVIATIONS

Btu/lb	Btu per pound
Cal/gm	Calorie per gram
cm	Centimeter
cfm	Cubic feet per minute
d	Day
ft	Foot (feet)
g (gm)	gram
gal	Gallon
gpd	Gallons per day
gpm	Gallons per minute
hr	Hour
in	Inch
kg	Kilogram
l	Liter
lb	Pound
m	Meter
mg/l	Milligrams per liter
ml	Milliliter
mm	Millimeter
MG	Million Gallons
MGD	Million gallons per day
min	Minute
min/rev (MPR)	Minutes per revolution
mo	Month
N/m^2	Newtons per square meter
Ns/m^2	Newtons-seconds per square meter
RPM	Revolutions per minute
sec (s)	Second
wt	Weight

SYMBOLS

BOD$_5$	Five Day Biochemical Oxygen Demand
CaCO$_3$	Calcium Carbonate
Ca(OH)$_2$	Calcium Hydroxide
CaO	Calcium Oxide
COD	Chemical Oxygen Demand
CST	Capillary Suction Time, sec
F:M	Food to Mass Ratio in Secondary Aeration System

$FeCl_3$	Ferric Chloride
$Fe(OH)_3$	Ferric Hydroxide
Hg (Inches Hg)	Mercury (Inches of Mercury Pressure-Gage)
MLSS	Mixed Liquor Suspended Solids
MLVSS	Mixed Liquor Volatile Suspended Solids
mm H_2O	Millimeters of Water Pressure-Gage
NH_3	Ammonia
NO_3	Nitrate
P	Total Phosphrous
pH	Hydrogen Ion Concentration
PO_4	Phosphate
psig	Pounds per square Inch-Gage Pressure
ΔP	Pressure Drop
Rp	Specific Resistance to Filtration, dimensionless (pressure)
Rv	Specific Resistance to Filtration, cm/g (vacuum)
SRT	Sludge Retention Time
TKN	Total Kjeldahl Nitrogen
TS	Total Solids
USDA	United States Department of Agriculture
VS	Volatile Solids
%	Percent
#	Pound
μ	Viscosity of water

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SECTION 1

BACKGROUND AND INTRODUCTION

The District of Columbia's Wastewater Treatment Plant at Blue Plains receives flow from the District of Columbia and from suburban Maryland and Virginia jurisdictions. Approximately two million residents produce an average daily flow of 1.06 Mm^3/day (280 MGD). Wastewater is treated by primary sedimentation and a secondary waste activated sludge process with chemical addition for solids capture and phosphorus removal. Sludge treatment is accomplished by two methods: gravity thickening and raw sludge dewatering with subsequent composting or trenching; or gravity thickening, anaerobic digestion, elutriation and dewatering with subsequent land spreading. (See Figure 1.)

An on-going expansion and upgrading of the plant, with completion scheduled for mid 1980, will add nitrification and multi-media filtration to the wastewater train. Sludge production will increase from its current level of approximately 150,000 kg/day dry solids (165 dry tons per day) to 340,000 kg/day solids (374 dry tons per day). To handle the additional sludge quantities, original plans had called for gravity thickening of primary sludge, air flotation thickening of all waste activated sludges, blending, vacuum filtration, and incineration. (See Figure 2.) All units except the incinerators have been installed. Because of the large amount of fuel oil which would be required to incinerate the vacuum filtered sludge cake, approval of the incinerators has been deferred by EPA pending further study. An initial study conducted by Camp, Dresser and McKee, Inc.[1] recommended a dual disposal system of composting and incineration and pointed out that if a high-solids sludge cake were produced, incineration could be accomplished with minimum quantities of auxiliary fuel. The study estimated that to incinerate 374 dry tons per day of an 18% vacuum filter cake, annual fuel oil usage would be approximately 16,000,000 gallons. Whereas, to incinerate a 35% solids cake, fuel usage would decrease to only 5,500,000 gallons per year. This 35% solids cake would be autocombustible and fuel would be required primarily for the afterburner to control toxic organics in the off-gas. Fuel usage in the multiple hearth furnace itself

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[1] Camp, Dresser & McKee, Inc. Alternative Sludge Disposal Systems for the District of Columbia Water Pollution Control Plant at Blue Plains. December, 1975.

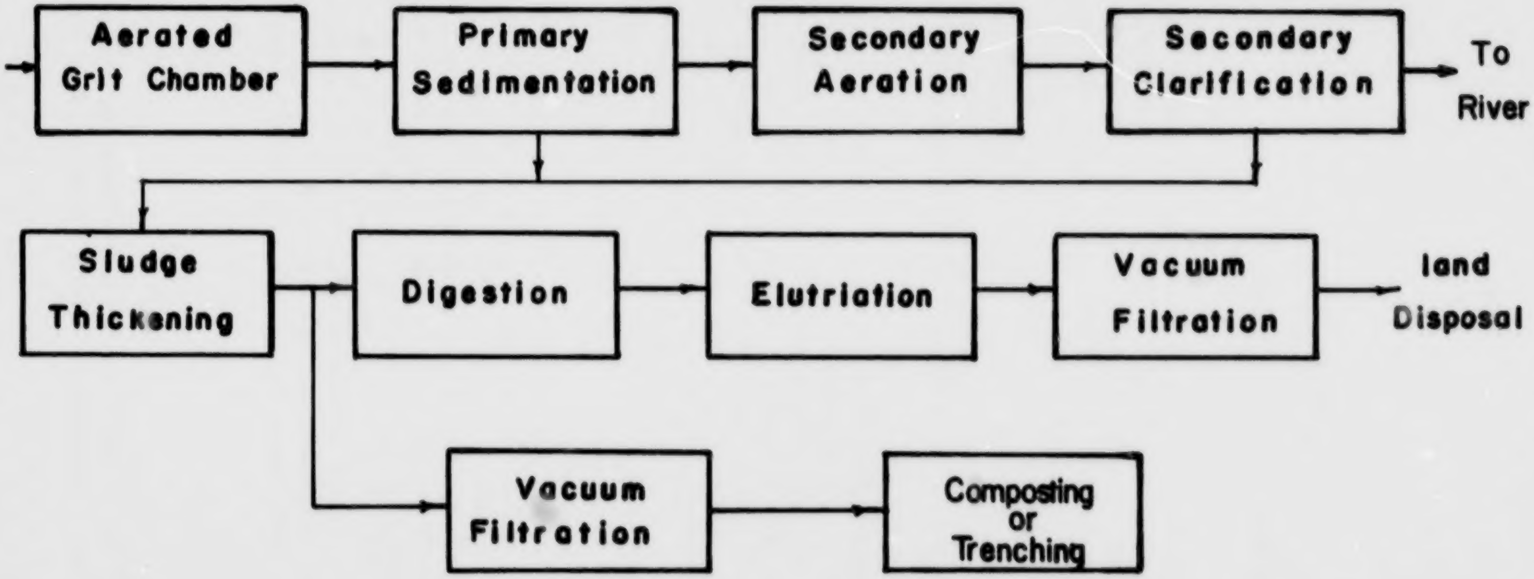


Figure 1. District of Columbia Wastewater Treatment Plant. Present facilities.

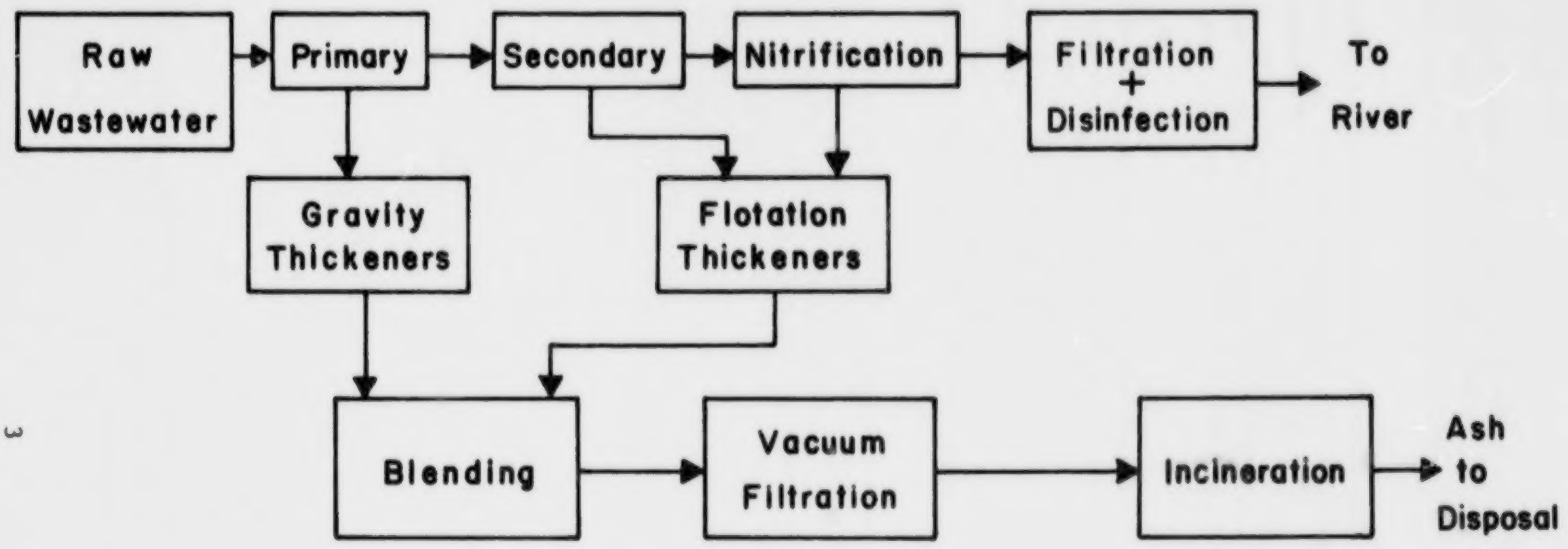


Figure 2. District of Columbia Wastewater Treatment Plant. Future facilities.

would be needed only for control of the position of the burning hearth. Camp, Dresser and McKee, therefore, recommended that the vacuum filters be replaced by filter presses so that an autocombustible sludge cake would be produced.

In mid-1976 the District of Columbia received approval from EPA to conduct a one year study of alternative dewatering devices for producing a high-solids sludge cake. The study was funded from two EPA sources. EPA Region III allowed expenses up to $186,992 as an addendum to an existing capital outlay project. EPA's Municipal Environmental Research Laboratory in Cincinnati, Ohio provided $49,693 and technical direction for the study.

The study was performed in the EPA-DC Pilot Plant using the existing equipment available in that facility as well as equipment provided by various manufacturers. A project engineer, a part-time chemical engineer, four sewage disposal plant operators, one chemist, and one mechanic conducted the entire study. The study period officially commenced on September 1, 1976 and ended on October 31, 1977; some preliminary test work actually began in April, 1976. The study was completed with expenses well below the budgeted amount.

The purpose of the study was to compare the operation of the various dewatering devices on selected ratios of waste activated to primary sludges. Because the plant had already purchased and installed 30 vacuum filter units, the District was interested only in evaluating devices that would produce significantly higher solids than the vacuum filters. Specifically, the District was interested in the results with a 2:1 ratio of waste activated-to-primary sludge solids. All units were operated in an attempt to produce an auto-combustible sludge cake. For sludges conditioned with inert chemicals, this requires a solids content of approximately 35%. The type of units tested and their suppliers included:

1. Vacuum Filter - Pilot model owned by EPA.
2. Vacuum Filter add-on devices - supplied by

    a) Envirotech Corporation, Salt Lake City, Utah.
    b) Parkson Corporation, Fort Lauderdale, Florida.
    c) Komline-Sanderson Corp., Peapack, New Jersey.

3. Belt Press - supplied by Parkson and Komline-Sanderson.
4. Filter Press - fixed volume @100 psig pressure supplied by Neptune-Nichols, Belle Mead, New Jersey.
5. Filter Press - fixed volume @225 psig pressure supplied by Passavant Corp., Birmingham, Alabama.
6. Filter Press - diaphragm type supplied by

    a) NGK Insulators, Ltd., Nagoya, Japan. Envirex Corporation has since purchased the rights to manufacture and market this press in the United States.

b) Dart Industries, Paramus, New Jersey.
c) Ingersoll-Rand Corp., Nashua, New Hampshire.

This report presents the results of testing the above dewatering units. Chemical conditioning with lime and ferric chloride was examined in detail to establish the requirements as a function of the ratio of waste activated-to-primary sludge solids. A correlation between bench-scale filterability tests and filter press performance was developed to monitor the conditioning step. Polymer conditioning was evaluated as an alternative to lime and ferric chloride conditioning.

Detailed design parameters for each of the filter presses were developed for a 2/1 secondary/primary sludge ratio. Comparison runs on these presses with the same batch of sludge provided valuable information on the advantages/disadvantages of each. Filter press cake was used in a variety of experiments to test cake shredding, incineration (solid waste furnace, multiple hearth furnace, and coal-fired boiler), and composting (static pile method).

The belt presses were used to provide design criteria for the thickened sludges. These presses were also modified to function as add-on units to further dewater vacuum filter cake.

Capital and operating costs and utility consumption are detailed for the dewatering units. Total disposal costs for dewatering plus both incineration and composting are also presented.

SECTION 2

SUMMARY AND CONCLUSIONS

Chemical Conditioning

1. The lime and ferric chloride dosages required to produce a filterable sludge varied with the percentage of waste activated sludge. Fibrous primary sludge filtered quite readily; waste activated sludge required greater quantities of conditioners and was more difficult to dewater. Generally, a 3/1 ratio of lime-to-ferric chloride was optimum for conditioning the Blue Plains sludge.

2. Laboratory tests showed that over-agitation of the conditioned sludge was detrimental to the filtration process. Floc deterioration with both time and high shear was a major factor in determining chemical requirements.

3. The addition of lime and ferric chloride to the sludge mixture increased the final dry weight of the filter cake by a corresponding amount. All of the iron and 80% of the calcium exited with the cake solids during filtration operations.

4. Bench-scale filterability tests were found to be useful when optimizing and controlling the lime and ferric chloride dosages.

5. Polymer conditioning of the 2/1 mixture of secondary-to-primary sludge was generally ineffectual. No single polymer was found which could adjust to the daily variations in the quality of sludge received from the primary and secondary treatment processes.

Filter Press-General

1. Each of the filter presses was capable of dewatering all sludge ratios in the range of 2.4-10% total feed solids to at least a 30% solids cake. The diaphragm press, however, was the only unit capable of dewatering the marginally conditioned sludges to the 35% solids required for an autocombustible cake.

2. Once a minimum chemical conditioning requirement of lime and ferric chloride for adequate dewatering was established, increases in filtration yields (up to 20%) were obtained by slight increases in chemical dosages.

3. In all the presses, suspended solids recovery in the filter cake was greater than 99%. The quantity of suspended solids in the filtrate was affected primarily by the type of filter cloth used and the degree of chemical conditioning.

4. The filter presses did not satisfactorily dewater polymer conditioned sludges.

5. The average specific resistance-to-filtration parameter was correlated directly with filter press yield.

Filter Press-Diaphragm Unit

1. On the average, the NGK press, using conditioning of 19.6% lime and 6.5% $FeCl_3$, dewatered the 2/1 secondary-to-primary sludge to a 38.7% solids cake with a yield of 2.39 $kg/hr/m^2$ (0.49 $lb/hr/ft^2$). The pumping pressure required to feed the press was always less than 7 kg/cm^2 (100 psig). The pumping cycle time averaged 17 minutes and was controlled by monitoring the total solids feed rate. A squeezing pressure of 15.0 kg/cm^2 (213 psig) was generally used. The squeezing cycle time (18 minutes) was controlled by filtering to a specified filtrate flow rate.

2. Equivalent results were obtained on the Ingersoll-Rand Lasta press. The full-scale yield for the unit, however, was somewhat higher at 2.93 $kg/hr/m^2$ (0.60 $lb/hr/ft^2$) for the 2/1 secondary-to-primary sludge.

3. Different filter cloths were tested on both the NGK and Lasta units. All gave acceptable filtrate quality but cloth life, resistance to abrasion, etc., could not be effectively evaluated in our study.

4. The cloth washing system in each of the presses also could not be adequately evaluated during the study. Maintenance of satisfactory cloth permeability by high-pressure sprays or acid washing is an area that generally requires more study.

Filter Press - Fixed Volume Unit

1. The high-pressure press (225 psig) had an average filtration yield of 1.51 $kg/hr/m^2$ (.31 $lb/hr/ft^2$) and required 62.3% more filtration area than the NGK diaphragm unit to produce equivalent results. The low-pressure press (100 psig) had an average full-scale yield of 1.07 $kg/hr/m^2$ (.22 $lb/hr/ft^2$) and needed 126.8% more filter area than the NGK diaphragm press to produce equivalent results.

2. Cycle time on the presses averaged 2-3 hours and was determined by filtering to a specified filtrate flow rate.

3. The cakes from the fixed volume presses always contained a dry outer section and a wetter inner core. This resulted in a substantial variation in the solids content across the cake.

Continuous Belt Press

1. Because of the highly variable sludge at Blue Plains, no polymer was found that could adjust to these variations and adequately condition the sludge at all times. The operation of the belt press, therefore, was not consistent.

2. With thickened sludge feeds, the press capacity, final cake solids, and polymer consumption were all affected by the percentage of waste-activated sludge. The unit performed best when dewatering high percentages of fibrous, primary sludge.

3. Suspended solids recovery in the filter cake averaged only 95%. Because of the stringent advanced waste treatment standards at Blue Plains, this level of recovery would be insufficient for continuous operation at this plant.

Vacuum Filter Retrofit Unit

1. The Envirotech Hi-Solids filter was discounted as an option for Blue Plains. It was capable of increasing the cake from a rotary vacuum filter to only 25% solids.

2. The use of the high-pressure section of the continuous belt press to further dewater the vacuum filter cake showed great promise. Cake solids of 35% were achieved in bench-scale work; however, demonstration of the system in a full-scale test was not successful because of problems with feeding the vacuum filtered cake to the press.

Filter Cake Processing

1. The filter press cake was composted with wood chips by the static-pile method. A good final product was produced with projected costs less than those for composting with vacuum filter cake.

2. Filter press cake with a solids content of at least 35% is considered a low-value fuel. It will burn in a multiple hearth incinerator without auxiliary fuel to produce an exit temperature of 800° F. It can also be co-burned with municipal refuse in a rocking-grate furnace. Because of the high ash content of the cake (up to 50%), however, it has been rejected as a fuel for a coal-fired boiler.

Economics

1. The belt press ($32.39 per ton) and the vacuum filter ($39.10 per ton) provide the lowest cost for dewatering.

2. Dewatering costs for each of the filter presses are nearly equal with unit costs of approximately $55.00 per ton.

3. Total disposal costs for filter pressing and incineration are approximately $88 per ton. This compares to the total cost for vacuum filtering and incineration at $130 per ton. Therefore, savings of nearly $4,000,000 per year for a 250 ton-per-day plant are possible by selecting filter presses for dewatering.

4. Total disposal costs for filter pressing and composting (including the cost of hauling the press cake 25 miles) are approximately $102 per ton. This compares to the total cost of vacuum filtering and composting (including hauling) of $155 per ton. Choosing a filter press rather than a vacuum filter, therefore, will result in annual savings of nearly $5,000,000 for a 250 ton-per-day plant.

SECTION 3

RECOMMENDATIONS

1. Filter presses should be installed at Blue Plains, designed to dewater the total quantity of sludge (average 374 dry tons per day) to be processed by either incineration or composting. This study showed that the diaphragm-type press offers the most flexibility and provides the best product. However, a final decision on the type of press to be utilized should be deferred until full-scale facilities of each type are inspected. Regardless of the type of unit chosen, a single large-scale unit should be purchased, installed and operated for several months to provide valuable design information prior to a large-scale committment of funds.

2. Additional test work should also be conducted to determine whether the specific resistance parameter can be used to successfully monitor and control the chemical conditioning process. As outlined in Section 8 (Process Design) a pilot-scale horizontal vacuum filter, adjusted to simulate the Buchner funnel filtration test, would be used for this purpose. When a pilot-unit becomes available, this work can be carried out in conjunction with existing vacuum filter operations.

SECTION 4

TYPE OF SLUDGE PROCESSED

During the study period, the wastewater treatment system included degritting, primary sedimentation, and a high-rate waste activated secondary process, with chemical addition for phosphorus removal. Table 1 shows average operating parameters of the system for FY 1977 (October, 1976 through September, 1977). The primary and secondary sludges are blended in a thickening operation and a portion of this undigested sludge is dewatered on vacuum filters.

Throughout the study period, the plant experienced continual operating problems because of an overloaded sludge processing system. Recycle loads, especially from gravity thickening, created operating difficulties in secondary. The recycle flows (only 5% of total flow) contributed 22% of the BOD loading and 25% of the suspended solids loading to the wastewater treatment train. This recycle problem, together with normal operating problems experienced with chemical addition outages and the rapidly changing biology in secondary, caused a highly variable sludge product. The combined sewer system in the District of Columbia also contributed to the problem; heavy rains washed large quantities of solids into the primary sludge thus changing the character of that product. As a result, the sludge dewaterability varied daily.

In setting up the study, the engineers attempted to simulate the conditions that would exist in the future full-scale plant. When the systems are completed, all primary sludge will be gravity thickened separately. The sludges from secondary and nitrification will be combined and air float thickened. All chemical precipitate will be included with the waste activated sludges. Backwash water and solids from the multi-media filters will be returned to secondary. Calculations of future sludge production show that the plant will produce an average ratio of 33% primary solids/67% waste activated solids.

The pilot plant had the capability to receive either blended thickened sludge from the plant or primary and secondary sludges separately. Initial test work with the plant thickened sludge gave good filtration results; however, the overloaded plant thickeners tended to wash out the fine solids. In order to get a more realistic product for dewatering, separate pilot-scale gravity thickeners for the primary and secondary sludges were placed in operation. A major variable for study was the dewaterability of various ratios of secondary to primary sludges. Consequently, the separate gravity thickening systems were used and the two sludges blended as necessary on a dry solids

TABLE 1. BLUE PLAINS WASTEWATER TREATMENT PLANT OPERATING PARAMETERS

Primary Treatment	Mean
Flow, Mm^3/day (MGD) including plant recycle	1.1 (292)
Influent suspended solids, mg/l	183
Influent BOD_5, mg/l	186
Detention time, hours	2.6
Surface Loading Rate, $m^3/day/m^2$ (gpd/ft^2)	42.3 (1039)
Suspended solids removal, %	45.0
BOD_5 removal, %	37.0
Sludge Production, kg/Mm^3(lb/MG)	78,960 (657.5)
Sludge wasted, % total solids	0.525
Sludge wasted, % volatile solids	72.4
Secondary Treatment	
Aeration Tank	
Flow, Mm^3/day (MGD)	1.1 (288)
Influent suspended solids, mg/l	102
Influent BOD_5, mg/l	121
Influent phosphorus, mg/l as P	6.2
MLSS/MLVSS concentrations, mg/l	1297/852
Detention time, hours	1.58
SRT, days	0.64
F:M, $days^{-1}$	1.51
Sedimentation Tank	
Detention time, hours	2.63
Surface loading rate, $m^3/day/m^2$ (gpd/ft^2)	33.1 (813)
Chemical Addition	
Ferric chloride, mg/l	23
Polymer (anionic), mg/l	0.20
Process Performance	
Suspended solids removal, %	72
BOD_5 removal, %	77
P removal, %	63
Sludge wasted, kg/Mm^3 (lb/MG)	118,410 (986)
Sludge wasted, % total solids	1.4
Sludge wasted, % volatile solids	66.1
Percent, biological solids/chemical solids	80/20
Fe content of waste sludge, %	10.0
Gravity Thickening	
Hydraulic loading rate, $m^3/day/m^2$ (gpd/ft^2)	29.9 (734)
Solids loading rate, $kg/day/m^2$, ($lbs/day/ft^2$)	118.8 (24.3)
Solids capture efficiency, %	77
Vacuum Filtration	
Feed, % total solids	7.0
Lime addition, % of feed solids	21.8
$FeCl_3$ addition, % of feed solids	7.5
Cake solids content, %	23.2
Filter yield, $kg/hr/m^2$ ($lb/hr/ft^2$)	15.3 (3.12)

basis. Air flotation thickening of the secondary sludge would have been the preferred method. Unfortunately, the logistics of running an air float thickener proved too cumbersome. Gravity thickening of the secondary sludge did produce the required 4 to 6% solids.

The primary sludge was delivered to the pilot plant and gravity thickened to 6 to 10% solids for use in the study. Normally, sludge was delivered continually from Monday morning to Friday and wasted as necessary to keep a sludge blanket in the thickener. The capacity of the thickener far exceeded the requirements for dewatering. The thickener was drained each Friday and fresh sludge started each Monday. Because the primary sludge at Blue Plains is very high in fiber content, a shredder was installed in the primary sludge delivery line (0.5% solids stream). It was used intermittently to keep the rags and trash from plugging the transfer pumps.

The secondary waste sludge was pumped directly from the secondary clarifiers to a gravity thickener in the pilot plant. This thickener was used primarily as a holding tank. Each evening, sludge was pumped in at a low flow rate. The flow was then cut off early in the morning and the contents allowed to thicken to 4-6% solids content. Sludge was used from this source during the day and the remaining contents drained each afternoon. Such an operation kept the sludge as fresh as possible. By severely limiting the overflow rate practically all the fines in the secondary sludge thickeners were captured.

A typical 2/1 secondary/primary sludge had the following characteristics

Percent solids	4 to 6
pH	6.2 to 6.8
Density, gm/cc (lb/gal)	1.006 (8.4)
Temperature, winter °C	8-15
Temperature, summer °C	25-30
% iron as Fe	7
% volatile solids	60-65

For all test runs, the sludges were blended in the following manner: Thickened secondary sludge (at 4-6% solids) was pumped with a Moyno pump to a calibrated mixing tank; the volume was measured, the sludge density measured, and a sample analyzed for percent solids on an O'Haus Moisture Balance. The quantity of dry solids in the tank was then calculated. The primary sludge was also analyzed for density and % solids, and the pounds of primary sludge calculated for a given volume. Based on the ratio of secondary to primary solids required, the volume of primary sludge was then Moyno pumped to the mixing tank. This method did give some experimental error; however, only approximate ratios were required for the type of tests run.

When plant thickened sludge was used for testing, it was pumped from the gravity thickeners to a tank truck (700 gallons) and then transported one-quarter mile to the pilot plant. A Moyno pump was used to pump the truck contents to the filter feed tank.

SECTION 5

SLUDGE CONDITIONING

Prior to either pressure or vacuum filtration, wastewater sludges must be chemically conditioned. A filter press will generally use ferric chloride and lime for conditioning. A vacuum filter will use either ferric chloride and lime or ferric chloride and polymer; belt presses will generally use polymer conditioning alone. For the purpose of this study, lime, ferric chloride and various polymers, either singly or in combination with one another, were examined for their suitability in conditioning the sludge. The study attempted to optimize each of these chemicals for each of the dewatering units. Other chemicals, such as aluminum chlorohydrate or ferrous sulfate, were not tested because they were either too costly or in short supply.

CONDITIONING WITH LIME AND FERRIC CHLORIDE

The lime used for the study was a bagged, pulverized, high calcium (94% CaO) quicklime. Lime dosages are reported as the weight of the lime as purchased. The ferric chloride used was purchased as a 30% by weight solution and was diluted as necessary. Results are reported on a 100% $FeCl_3$ basis. Table 2 gives the specifications for the lime and ferric chloride. Percent chemicals (either lime or $FeCl_3$) are calculated as:

$$\frac{\text{lbs dry weight of chemical}}{\text{lbs dry incoming sludge solids}} \times 100 = \text{\% chemical}$$

Material Balance Tests

The addition of lime to the thickened sludge stream is expected to form calcium carbonate (insoluble) and calcium hydroxide (soluble). The quantities normally required will raise the pH of the solution to 11.0 or above. Ferric chloride reacts at this high pH to form the insoluble ferric hydroxide ($Fe(OH)_3$). The disposition of these metal ions was determined with a material balance test on a Buchner funnel. At three different chemical dosages, approximately 225 ml of conditioned sludge was filtered; the feed, cake and filtrate were all analyzed for calcium (Ca) and iron (Fe) content. The Ca and Fe determinations were made with an Atomic Absorption Spectrophotometer. Table 3 shows the test results. Note that the weight of Ca and Fe in the feed do not balance exactly with the Ca and Fe in the filtrate and cake. However, the tests are useful in that they show that approximately 80% of the calcium and 100% of the iron will exit with the cake. The remainder of the calcium stays with the filtrate, probably as calcium hydroxide.

TABLE 2. CHEMICAL SPECIFICATIONS

Lime	
Type	Pulverized CaO from Warner Co., Bellefonte, Pa.
CaO	94.84%
Available CaO	92.1%
Size	100% will pass a 100-mesh sieve 1.95 micron average particle diameter
Concentration (as used)	1 lb/gallon (11.3% by weight)
Specific gravity	1.06
Ferric Chloride	
Type	Liquid @ 30% by weight from DuPont titanium dioxide manufacture
Concentration (as used)	1 lb/gallon (10.9% by weight)
Specific gravity	1.103

Additional tests were run on the Buchner funnel for the purpose of performing a total solids balance (determined by evaporating the samples to dryness). These tests, at different chemical dosages, were carefully run to determine the increase in dry solids due to the addition of lime and ferric chloride. Thickened sludge (primary and waste activated) was conditioned and approximately 200 ml were filtered. Weights and total solids of the unconditioned sludge, the conditioned sludge, the cake and the filtrate were measured. The results are shown in Table 4. The tests show that as chemicals are added, the weight of solids actually increases above the initial weight of sludge plus chemicals added. This is most likely due to $CaCO_3$ formation. The total solids in the final cake, however, match very closely with the initial weight of sludge plus the weight of lime and ferric chloride added. The conclusion we reached from these tests was that both the lime and ferric chloride weights must be accounted for in the filter cakes off either a vacuum filter or filter press. In all calculations we therefore assumed that for every pound of lime and ferric chloride added for conditioning the final dry weight of the filter cake also increased by an identical amount.

Important Considerations In Conditioning

A considerable amount of trial-and-error work on the filter presses and the bench studies showed the following:

1. For Blue Plains sludge, the minimum amount of $FeCl_3$ needed for conditioning was approximately 5% by weight of sludge solids.

TABLE 3. MATERIAL BALANCE TESTS FOR Ca AND Fe

% CHEMICAL lime/FeCl$_3$	FEED SLUDGE gms Ca	Fe	Solids	FILTRATE gms Ca	Fe	CAKE gms Ca	Fe	Solids
15.6/5.4	1.107	.770	15.7	.232	.00012	.865	.806	15.1
20.8/6.9	1.450	.824	16.5	.277	.00014	1.106	.788	15.6
26.1/8.7	1.604	.837	16.3	.389	.00014	1.294	.892	15.7

TABLE 4. MATERIAL BALANCE TESTS FOR TOTAL SOLIDS

UNCONDITIONED SLUDGE gms +	LIME gms +	FeCl3 gms =	TOTAL SOLIDS COMPUTED IN FEED gms	TOTAL SOLIDS MEASURED IN CONDITIONED SLUDGE gms	TOTAL SOLIDS MEASURED IN CAKE gms	TOTAL SOLIDS MEASURED IN FILTRATE gms
25.68	4.08	1.36	31.12	32.67	31.32	1.96
23.93	5.09	1.70	30.72	32.67	30.80	2.20
23.51	6.30	2.08	31.89	33.77	32.00	2.50
20.36	3.07	1.02	24.45	25.67	23.95	1.56
19.74	4.06	1.36	25.16	26.69	24.71	1.88

2. Three parts of lime per part of $FeCl_3$ worked all the time. However, the optimum lime:$FeCl_3$ ratio could be in the range from 2:1 to 4:1.

3. The $FeCl_3$ was always added first and allowed to mix thoroughly before adding the lime. The $FeCl_3$, however, forms a weak floc which can be easily broken up by too vigorous mixing; consequently care has to be exercised during the mixing.

4. After the lime has been thoroughly mixed in, the sludge should be filtered as soon as possible. The following tests were made to assess floc deterioration. Specific resistance tests were run with a Buchner funnel and the results reported as Rv. For good filtration on the pilot-scale NGK filter press the Rv value should be less than 27×10^{10} cm/gm. (See Appendix C for the Rv procedure). These laboratory tests were conducted in a 1500 ml beaker with a single paddle stirrer (1" high x 3" wide). Maximum speed of the stirrer was 100 RPM. Sludge was added to the beaker with the mixer at 100 RPM. $FeCl_3$ (6.2%) was added and mixed in thoroughly (approximately 5 to 6 minutes). Lime (18.6%) was added and mixed in (approximately 5 to 6 minutes). When a visual check showed that the chemicals were well dispersed, this was called time=0. At various time intervals samples were grabbed and the Buchner funnel test made to determine Rv.

   In Run #1, the mixer was allowed to operate at 100 RPM for the duration of the test. This run showed that the specific resistance increased rather quickly:

Time (min)	Rv (cm/g)
0	17.5×10^{10}
10	94.8×10^{10}
20	125×10^{10}

   In Run #2, the mixer also ran at 100 RPM for the entire test. By visual examination, the operator picked the time at which the sludge appeared to change. This test showed that the breakdown occurred in approximately five minutes:

Time (min)	Rv (cm/g)	
0	23.1×10^{10}	
5	76.1×10^{10}	sludge appeared milky
11	98.9×10^{10}	

   In Run #3, the mixer was slowed to 20 RPM after the lime and $FeCl_3$ had been mixed in thoroughly. This slow speed was barely adequate to keep the sludge mixed. A longer time was needed before noticeable floc deterioration:

Time (min)	Rv (cm/g)	
0	11.8×10^{10}	
10	27.3×10^{10}	
20	43.5×10^{10}	floc began to deteriorate
30	93.2×10^{10}	

The above tests confirmed the visual observations made throughout the entire test period; the same deterioration of sludge floc was observed many times in the NGK mixing tank. Marginally conditioned sludges were especially susceptible to over mixing or too-long storage times. If, however, the sludges were conditioned well above the marginal level, this rapid deterioration was less pronounced. Consequently, it should be noted that lime and $FeCl_3$ usage can be minimized by proper design and operation of the conditioning system.

Chemical Requirement vs. Secondary/Primary Sludge Ratio

A test in a Buchner funnel was used to show the effect of the ratio of secondary/primary sludge on chemical dosage requirements. This test was made for seven different sludge ratios. For each ratio, the sludges were blended in the proper proportions and then conditioned with lime and ferric chloride. In all cases, a 3:1 weight ratio of lime: $FeCl_3$ was used. The dosage was considered to be optimum if the sludge could be filtered down to a good cake in less than 3-4 minutes. The results of this test are shown in Figure 3. It should be noted that this graph shows only a trend, rather than absolute chemical requirements. The Blue Plains sludge varies to the extent that these results would not be duplicated if the test was repeated on a subsequent day. This trend, however, is exactly what was found with all the filter press runs. Generally, primary sludge, because of its high fiber content, filters quite readily with only low chemical requirements. Secondary sludge which is composed of small biological solids is more difficult to condition and filter. As the percentage of secondary sludge increases, the chemical requirements also increase. If enough lime and $FeCl_3$ are added, though, the sludge can always be made to dewater.

CONDITIONING WITH POLYMER

This topic will be discussed under each of the dewatering unit sections.

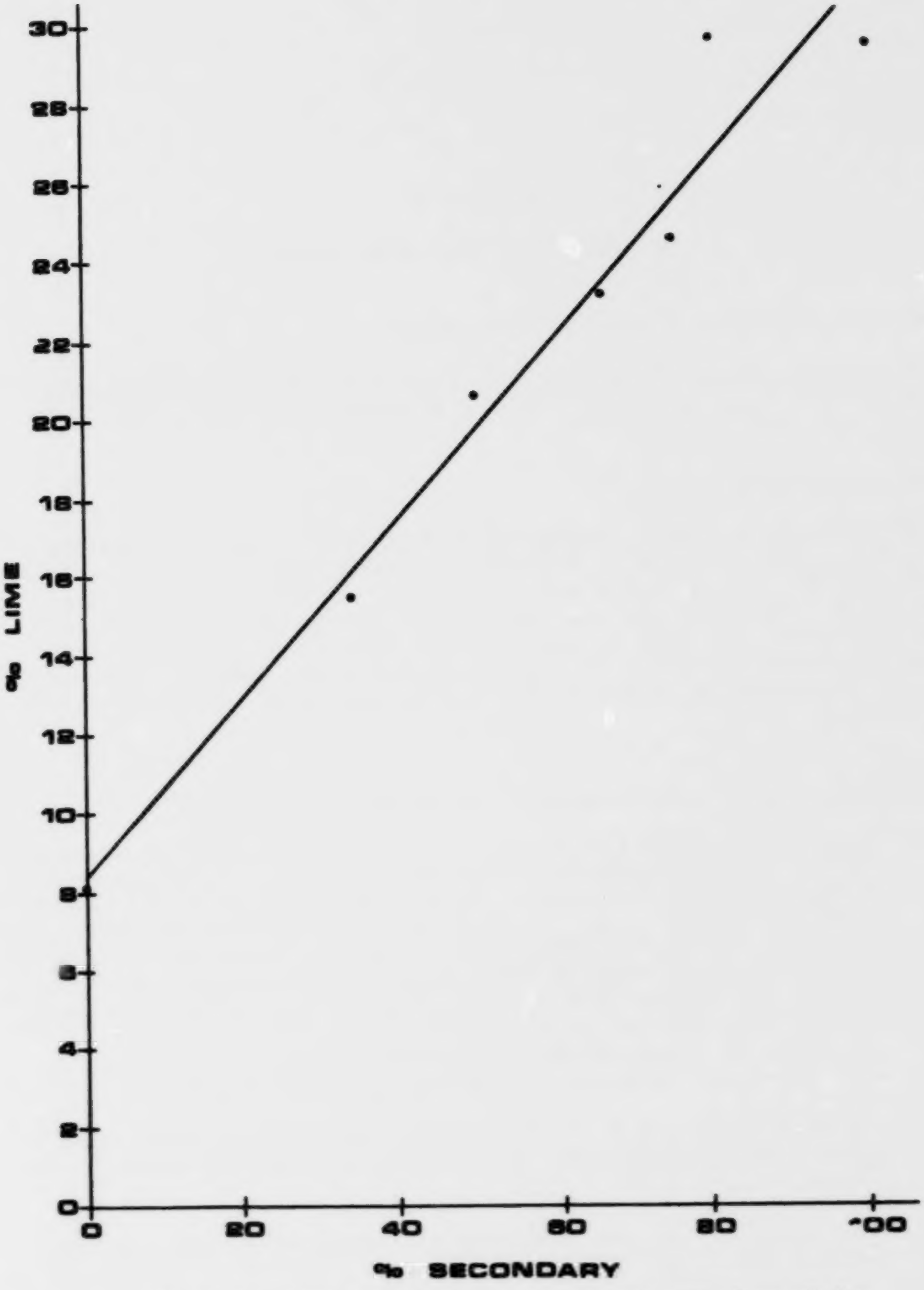


Figure 3. Lime requirements vs percent secondary sludge.

SECTION 6

TEST RESULTS - DEWATERING UNITS

DIAPHRAGM FILTER PRESS (VARIABLE VOLUME PRESS)

The diaphragm-type filter press, a relatively new innovation in the wastewater treatment industry in the United States, was tested most extensively during the study.

NGK Diaphragm Press

NGK Insulators, Nagoya, Japan, has been manufacturing and marketing a diaphragm-type filter press in Japan for several years. A pilot-scale model of their NR-PF-II filter press was provided to the District for the duration of the study. This unit was used not only to provide design parameters for a diaphragm press, but also to study several other factors associated with any sludge dewatering operation. Envirex Corporation, Waukesha, Wisconsin, has since purchased the rights to manufacture and market this press in the United States.

Facilities--

The filter press system included the following equipment:

1. Press - 5.8 m^2 (62.4 ft^2) filtration area. Contained six chambers, with twelve 800 mm (31.5 inches) square plates. Spacing between plates was 25 mm (1.0 inch). Every other plate was equipped with rubber diaphragms. As is typical of filter presses, the surface of the plate behind the filter cloth resembles the surface of a waffle iron, to allow removal of filtrate that passes through the filter cloth. The surface of the rubber diaphragm in contact with the filter cloth also has a raised grid pattern for this purpose. The press was equipped with a hydraulic closing mechanism and an overhead cloth vibrating and washing unit. See Figure 4.

2. Sludge mix tank - 1.0 m^3 (264 gallon) tank, with variable-speed mixer, equipped with three turbine-wing type agitator blades. See Figure 5.

3. Pump assembly - sludge feed pump, squeezing water pump, and cloth washing pump. See Figure 6.

    a. Feed pump - a diaphragm-type piston pump rated at 100 l/min (26 gpm) and pressures up to 7 kg/cm^2 (100 psig).

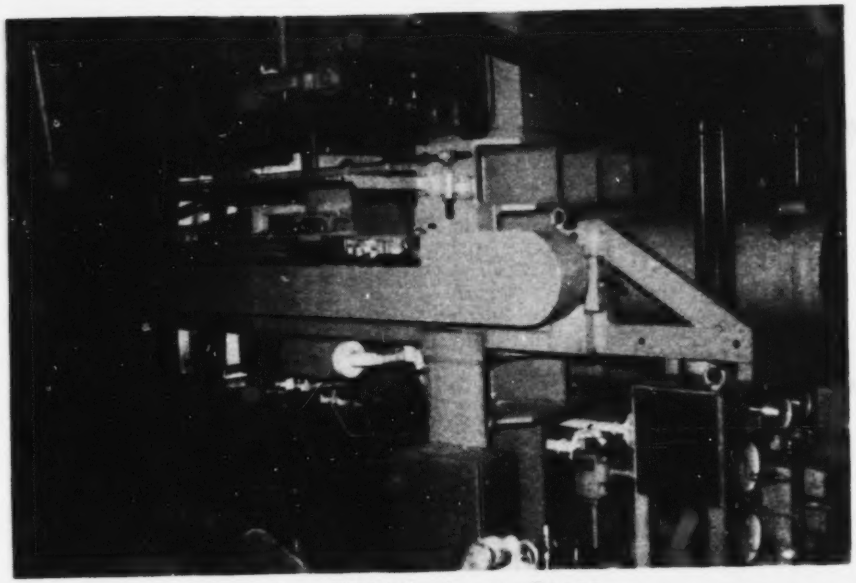


Figure 4. NGK Diaphragm Press.

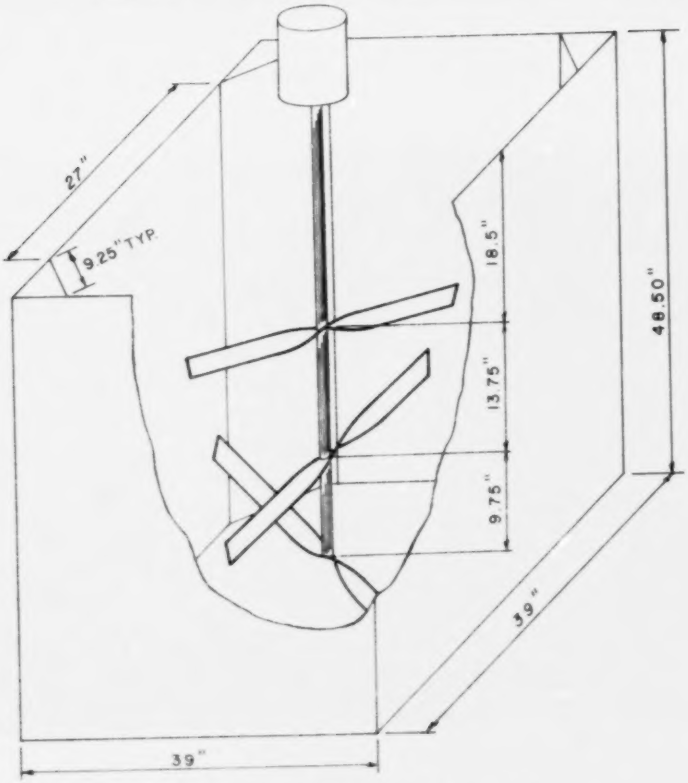


Figure 5. NGK sludge mix tank.

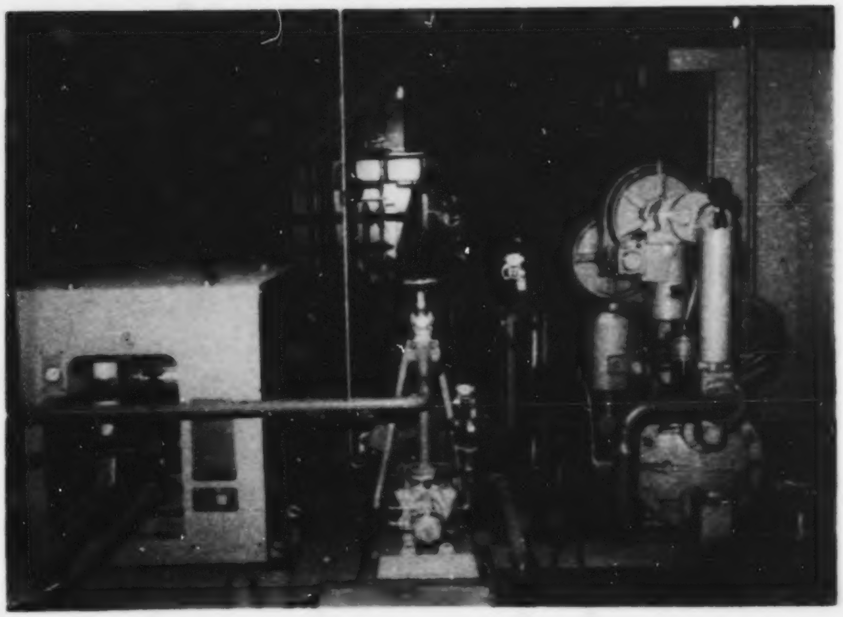


Figure 6. NGK pump assembly.

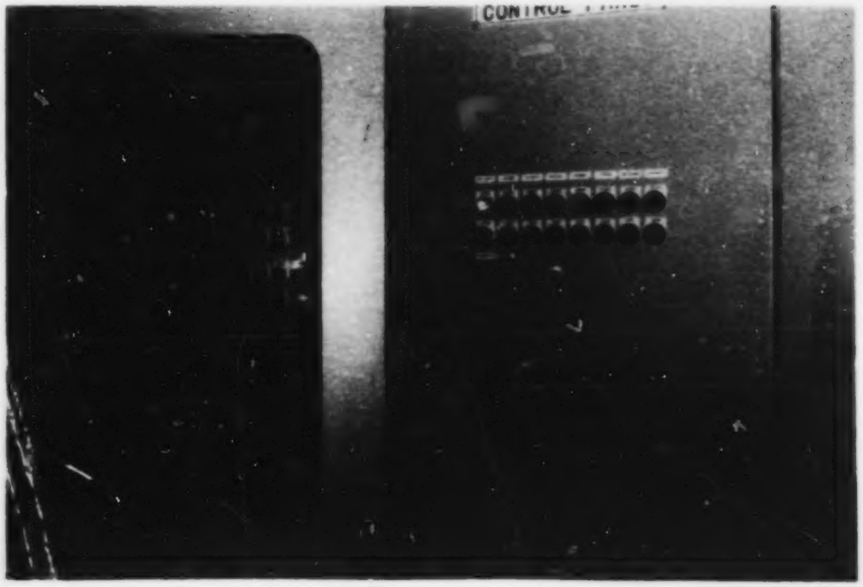


Figure 7. NGK control panel.

   b. Squeezing water pump - a multi-stage turbine pump rated at 60 1/min (15.8 gpm) and pressures up to 17.9 kg/cm^2 (255 psig).

   c. Cloth washing pump - plunger-type pump rated at 92 1/min (24 gpm) and pressures up to 70 kg/cm^2 (1000 psig).

4. Water storage tank - 500 1 (132 gallons).

5. Air compressors - with receivers (two), each rated at 7 kg/cm^2 (100 psig) for operating ball valves and core blowing.

6. Control panel - with relays, timers, etc. Allowed either a totally automatic or manual mode of operation. See Figure 7.

The District provided the following equipment to complete the system.

1. Primary sludge thickener - total volume of 28.4 m^3 (7500 gal) and overflow surface area of 8.9 m^2 (96 ft^2).

2. Secondary sludge thickener - total volume of 20.8 m^3 (5500 gal) and overflow surface area of 6.0 m^2 (65 ft^2).

3. Moyno transfer pumps - (two) each rated at 37.8 1/min (10 gpm).

4. Lime slurry makeup and storage tank - 757 1 (200 gal) with agitator.

5. Ferric chloride makeup and storage tank - 378 1 (100 gal).

6. Batch feed tanks - ferric chloride tank, 37.8 1 (10 gal); lime tank, 56.8 1 (15 gal).

7. Calibrated filtrate collection tanks - 378 1 (100 gal) and 56.8 1 (15 gal).

Operation--

A complete cycle for the NGK filter press included pumping, squeezing, and cake discharge operations. A typical cycle was as follows. See Figure 8.

Primary and secondary sludges were pumped into the sludge mix tank at the desired test ratio. Solids content of the mix was measured and the chemical dosage computed as a percentage of dry sludge solids. $FeCl_3$ (usually 5-10% by weight of dry sludge solids) was added by gravity and mixed in at an agitator speed of approximately 95 RPM. Lime (usually 15-30% by weight of dry sludge solids) was also added by gravity and mixed in at 95 RPM. After visual examination showed the chemicals to be well mixed (about 10-15 minutes), the mixer was slowed to a speed just sufficient to prevent stratification (approximatly 28 RPM). The filter press was closed by actuating the hydraulic unit, which held a constant pressure of 200 kg/cm^2 (2844 psig) on the plates during the entire press cycle. The filtration cycle began when the sludge feed pump was started. Pumping time was normally 10 to 20 minutes, allowing a sludge feed of 227 to 303 liters (50 to 80 gal). Figure 9

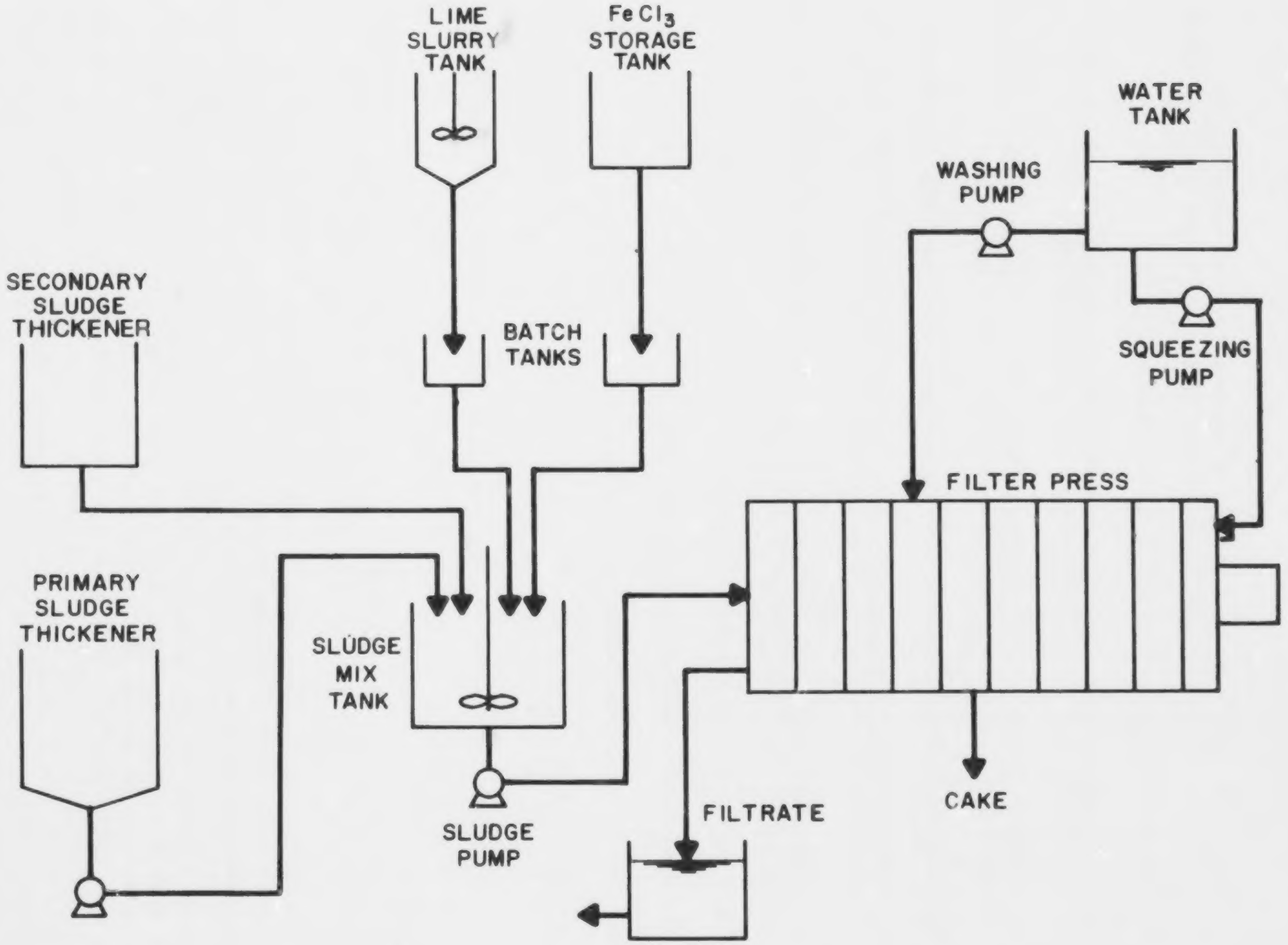
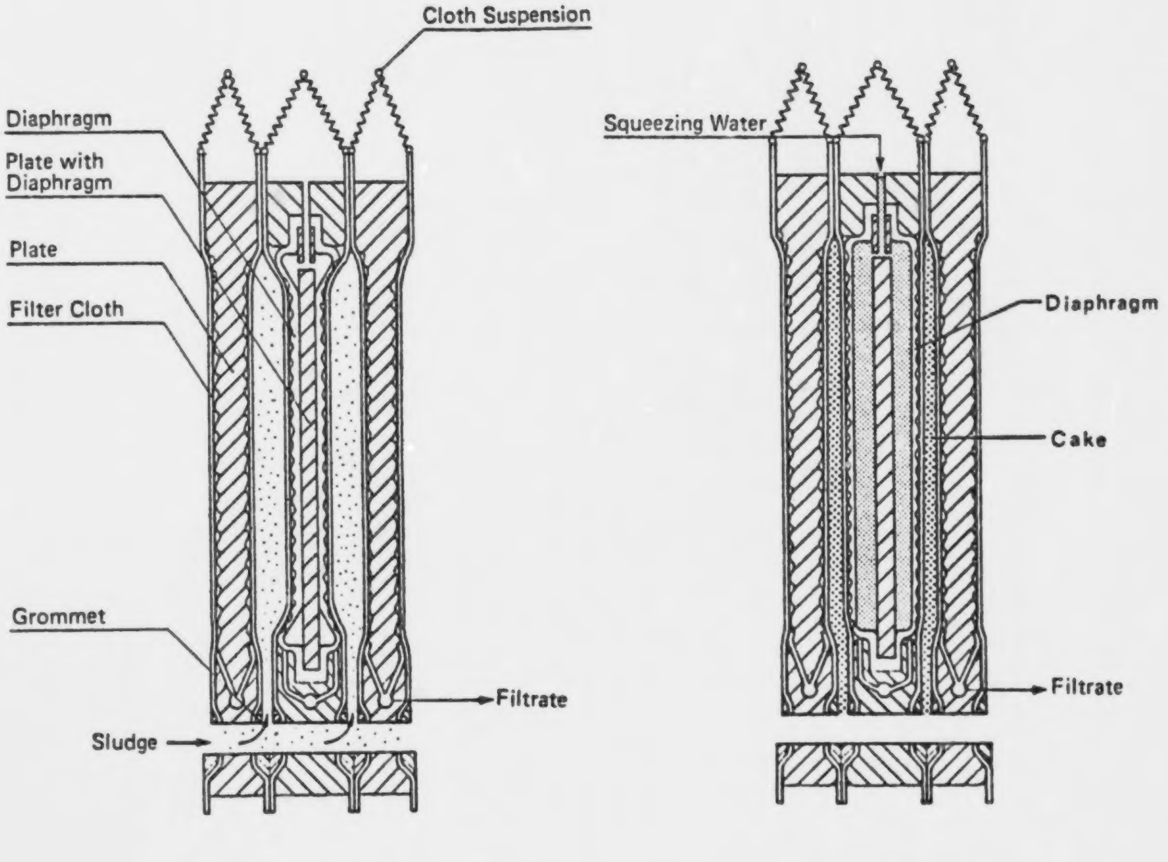


Figure 8. Process schematic for NGK diaphragm press.



FILTRATION

SQUEEZING

Figure 9. Schematic of filtration and squeezing in diaphragm press.

shows a detailed view of the filtration operation. Sludge is fed through a bottom feed port into the empty chamber. Filtered water passes through both cloths to collection ports on the ends of the plates. At the end of the pumping cycle, the squeezing pump was started immediately to pressurize the diaphragm. Squeezing time was usually 10-25 minutes at a pressure of 15 kg/cm^2 (213 psig). Figure 9 also shows how the pressurized water expands the rubber diaphragm behind one of the cloths in each chamber. The cake is compressed to approximately half its original thickness as the filtrate passes through both cloths.

At the end of the squeezing cycle, the sludge feed lines and filtrate lines were blown out with pressurized air. The press ram was opened and the plate shifter carriage moved two plates into position for cake discharging. The overhead vibrating unit subsequently positioned itself over these two plates, lowered its two vibrating shoes onto the cloth support bars, and shook the four cloths with an eccentric cam action, thereby discharging two cakes. The shifter carriage then moved another two plates and the process repeated automatically. The pilot unit discharged six cakes, each measuring 686 mm (27 inches) square and approximately 13 mm (0.5 inch) thick. At the end of the discharge cycle, the shifter carriage, the vibrating unit and the plates all moved back into position, ready for another run. At this point, the cloth washing cycle was initiated when required. The cycle was also completely automatic and similar in operation to that of the cake discharge. The overhead vibrating and wash unit was equipped with two spray bars, which washed four cloths at one time. The cloths were washed each morning and evening, and depending on the type of tests, after each run. Figure 10 shows detailed views of both the cake discharge and washing operations. The cloths are attached to the plates at the bottom but are suspended from springs at the top. The cloth moves away from the top of the plate to facilitate both the discharging and washing operations.

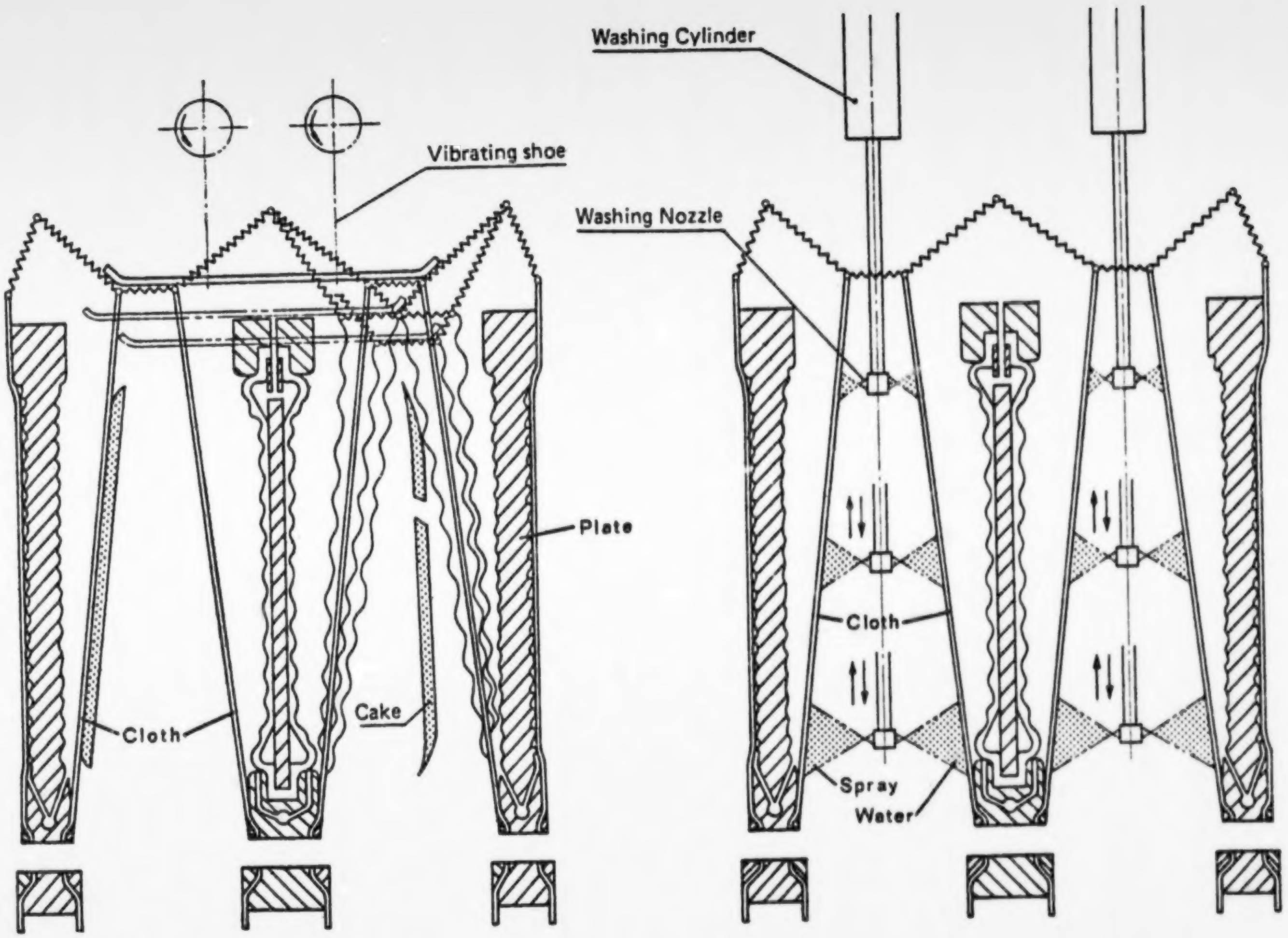
Data sheets 1 through 4 in Appendix B were used in recording data for the test runs. Raw data was recorded on sheets 1 through 3 and results summarized on data sheet 4. The data sheets are filled out for a typical set of runs, with calculations detailed in an accompanying explanation.

Test Data to Establish Design Parameters--

In order to develop design parameters for the dewatering of a 5% solids sludge to produce a 35% solids cake on a diaphragm filter press, the following parameters should be optimized:

1. Chemical requirements
2. Feed pump pressure
3. Pumping time
4. Squeezing pressure
5. Squeezing time
6. Filtrate quality
7. Filter cloth selection
8. Filter yield--as a function of all of the above.

Because of the constantly changing filtration characteristics of the sludge, it was extremely difficult to compare test results from one day



CAKE DISCHARGE

CLOTH WASHING

Figure 10. Schematic of discharge and washing in diaphragm press.

with those of another. Consequently, it became necessary to try to collect enough data for a good comparison study of these parameters from the same batch of sludge, preferably in one days' time. The following sections discuss the methods used to optimize each of the parameters. The section on filter yield discusses average overall results for the 2/1 secondary/primary sludge.

Chemical requirements--Generally, there are three possible ranges of chemical conditioning:

1. The level of lime/$FeCl_3$ dosages below which no dewatering will occur. It is obvious that this level must be defined and appropriate measures taken to insure that all sludges are conditioned above it. Adding underconditioned sludge to a press will cause many wasted manhours in cloth rejuvenation, either by high pressure spray washing or acid washing.

2. The level of chemical dosages in the optimum range where good filtration will occur. Higher chemical dosages within this range give slightly higher cake solids and filter yields. At a particular installation, the operator can choose to operate at the upper or lower end of this range depending on sludge quantities to be filtered. Obviously, cost savings in chemical will inspire the operator to stay in the lower end of the range as much as possible.

3. The level of chemical dosages at very high lime/$FeCl_3$ addition where the chemicals are overdosed but dewatering readily occurs. This is a very safe level for operation, with very little chance of press failure but chemical costs are extremely high. The increased quantities of inert solids in the final cake also result in decreased yields and could cause problems in further processing.

Table 5 shows the effect of varying chemical dosages on the major filter press parameters, final cake solids content and filter yield. Full-scale yield is defined as the weight of sludge solids per square meter of filtration area per hour. The total cycle time used in calculating this full-scale yield includes the pumping and squeezing times plus 19 minutes mechanical turnaround time (based on the manufacturer's recommendation for their largest press). See Appendix B, Explanation of Data Sheet 4. The runs on 2/18, 2/23, 3/4, 7/12 and 6/23 clearly show levels at which the sludge will not dewater. With all of these poorly conditioned runs, the filter cloths required considerable cleaning. In cases where the conditioning was adequate for dewatering, higher chemical dosages generally gave either higher cake solids and/or higher yields. Examination of the data shows that on the average, an increase of 5 percentage points of lime and 2 of $FeCl_3$ can give a 20% increase in full-scale yield. But, if the objective is to obtain a specific cake solids content, for example 35%, at minmum chemical addition, a point is reached beyond which chemical addition is wasteful. The runs on 3/10 show that no benefit in yield is gained by increasing the chemical dosage above 22.8% lime/6.7% $FeCl_3$. In fact, the very high dosages (45.5% lime/13.3% $FeCl_3$) actually showed a decreased yield because of the quantity of inert chemicals in the final sludge cake.

Chemical dosage was, in summary, largely a function of the sludge characteristics existing at the time. Establishing the proper dosage is a trial-and-error procedure that must be performed on each batch of sludge to be filtered. A simple Buchner funnel test and the use of the Capillary Suction Time meter can aid in defining this dosage. These methods are discussed further in the section on specific resistance tests.

Feed pump pressure--The sludge feed pump supplied with the NGK press was capable of delivering pressures from 3 to 7 kg/cm^2 (43 to 100 psig). Numerous tests to optimize the terminal pump pressure were inconclusive. The more difficult sludges generally would give higher filter yields if the pump pressures were above 5 kg/cm^2 (71 psig). The easier to filter sludges, such as those with high primary ratios, could be handled with lower pressures of 3 kg/cm^2 (43 psig). In large installations, optimization of pump pressure should be done under continuous operating conditions, while considering filter yield, chemical conditioning requirements, and especially, filter cloth life.

Pumping time--With a diaphragm press, the pumping as well as the squeezing cycle time must be optimized so that the filter yield will be maximized. The automatic control system supplied with the press provided the option of operating with a preset pumping time for each cycle. This mode of operation is best, however, only if the sludge filterability does not change, i.e., if a constant lime/$FeCl_3$ dosage gives consistent results on the filter press. With the Blue Plains sludge, this was not the case. Sludge filterability and, hence, required chemical dosages varied almost daily. With some sludges a pumping time of 5 minutes was sufficient; with others, 25 minutes was best. Figure 11 shows the variation of total feed volume with time in several press runs with different levels of conditioning. Data for these runs of 3/4/77 are shown in Table 5. For runs 1 and 2 in which the sludge was well conditioned, the feed rate remained quite high (over 5 gpm) until the ninth minute. After that time, the slopes began to flatten out as the resistance to filtration started to increase. In run 3, a poorly conditioned sludge, the feed rate dropped off and the resistance to filtration began to increase after only the fourth minute.

In a diaphragm-type press, the pumping cycle is used primarily for adding filterable solids to the press, and the pumping cycle time should be optimized to this end. For example, in Table 6, several runs are shown for 4/1/77 and 4/6/77 in which successively longer pump times were used. In each case, as the pump cycle was extended, a greater quantity of solids was added to the press (as evidenced by cake dry weights). Notice, though, that a corresponding increase in yield was not obtained. The key to optimizing the pumping cycle, i.e. to obtain the maximum yield for maximum solids input to the press, lies in knowing the solids addition rate for each successive minute of pumping. Once this rate drops below the expected average solids yield on the press (kg/hr), then the pumping cycle should be terminated. Determination of this rate was made for a generalized sludge feed to the NGK press and was correlated to a terminal sludge volume rate.

From early test work, we established an average rate of 2.4 kg total solids/hr/m^2 (0.5 lb/hr/ft^2) as a reasonable production rate for a full-scale

TABLE 5. NGK FILTER PERFORMANCE VS. CHEMICAL CONDITIONING

DATE	RATIO SEC/PRIM	% CHEMICALS LIME/$FeCl_3$	CYCLE TIME (min) PUMP/SQUEEZE	% CAKE SOLIDS	FULL-SCALE YIELD $kg/m^2/hr$	CAKE DISCHARGE
2-18-77	1/1	26.4/7.8	20/10	35.5	3.14	excellent
2-18-77	1/1	17.8/5.0	20/10	23.6	2.06	poor (wet)
2-23-77	1/1	32.0/9.4	20/10	34.6	2.69	excellent
2-23-77	1/1	18.3/5.4	20/10	23.2	1.76	poor (wet)
3-4-77	1/1	29.4/8.6	16/8	38.4	2.94	excellent
3-4-77	1/1	27.5/8.1	16/8	31.6	3.57	excellent
3-4-77	1/1	14.2/4.1	16/8	24.1	2.87	poor (wet)
3-10-77	1/1	45.5/13.3	16/10	39.6	2.85	excellent
3-10-77	1/1	34.3/10.0	16/10	38.5	3.25	excellent
3-10-77	1/1	22.8/6.7	16/10	36.0	3.27	excellent
3-10-77	1/1	17.1/5.0	16/10	31.6	2.51	good
7-8-77	1/1	18.9/6.4	16/14	44.0	3.80	excellent
7-8-77	1/1	14.0/4.6	16/16	41.2	3.46	excellent
7-12-77	1/1	16.8/5.6	14/19	41.8	2.76	excellent
7-12-77	1/1	14.4/4.9	12/23	39.9	2.22	excellent
7-12-77	1/1	12.2/4.1	11/23	29.0	1.44	poor
4-29-77	2/1	25.6/8.4	21/19	39.8	2.09	excellent
4-29-77	2/1	19.5/6.5	20/19	35.7	1.84	excellent
4-29-77	2/1	15.5/5.2	18/25	35.1	1.49	excellent
5-19-77	2/1	18.6/6.2	15/15	38.0	2.78	excellent
5-19-77	2/1	13.8/4.6	13/26	35.5	1.71	excellent

TABLE 5.

DATE	RATIO SEC/PRIM	% CHEMICALS LIME/$FeCl_3$	CYCLE TIME (min) PUMP/SQUEEZE	% CAKE SOLIDS	FULL-SCALE YIELD $kg/m^2/hr$	CAKE DISCHARGE
5-3-77	2/1	18.6/6.2	18/19	38.0	1.97	excellent
5-3-77	2/1	17.4/5.8	9/24	38.1	1.30	good (thin)
5-3-77	2/1	13.7/4.5	6/20	37.5	1.05	good (thin)
5-26-77	2/1	20.3/6.6	18/19	36.1	2.00	excellent
5-26-77	2/1	15.7/4.8	16/19	29.3	1.71	excellent
6-23-77	2/1	27.0/9.0	16/18	38.4	2.07	excellent
6-23-77	2/1	24.7/8.3	15/18	39.2	1.95	excellent
6-23-77	2/1	17.2/5.9	11/18	29.4	1.47	poor (wet)
8-19-77	2/1	21.5/7.2	20/16	42.4	3.19	excellent
8-19-77	2/1	14.2/4.8	18/21	36.0	2.12	excellent
10-18-77	2/1	20.1/6.6	14/15	41.0	3.18	excellent
10-18-77	2/1	14.8/5.0	17/18	35.5	2.47	excellent
6-30-77	3/1	24.6/8.3	19/13	40.0	2.97	excellent
6-30-77	3/1	19.9/6.8	20/18	39.6	2.47	excellent
6-30-77	3/1	15.7/5.4	18/21	40.4	2.23	excellent
9-15-77	3/1	26.8/8.9	24/21	36.0	1.96	excellent
9-15-77	3/1	21.3/7.1	23/21	31.2	1.63	good
11-1-77	1/0	25.2/8.3	17/16	38.5	2.67	excellent
11-1-77	1/0	19.7/6.6	17/17	36.8	2.73	excellent
11-1-77	1/0	14.8/5.0	13/21	34.5	2.01	good

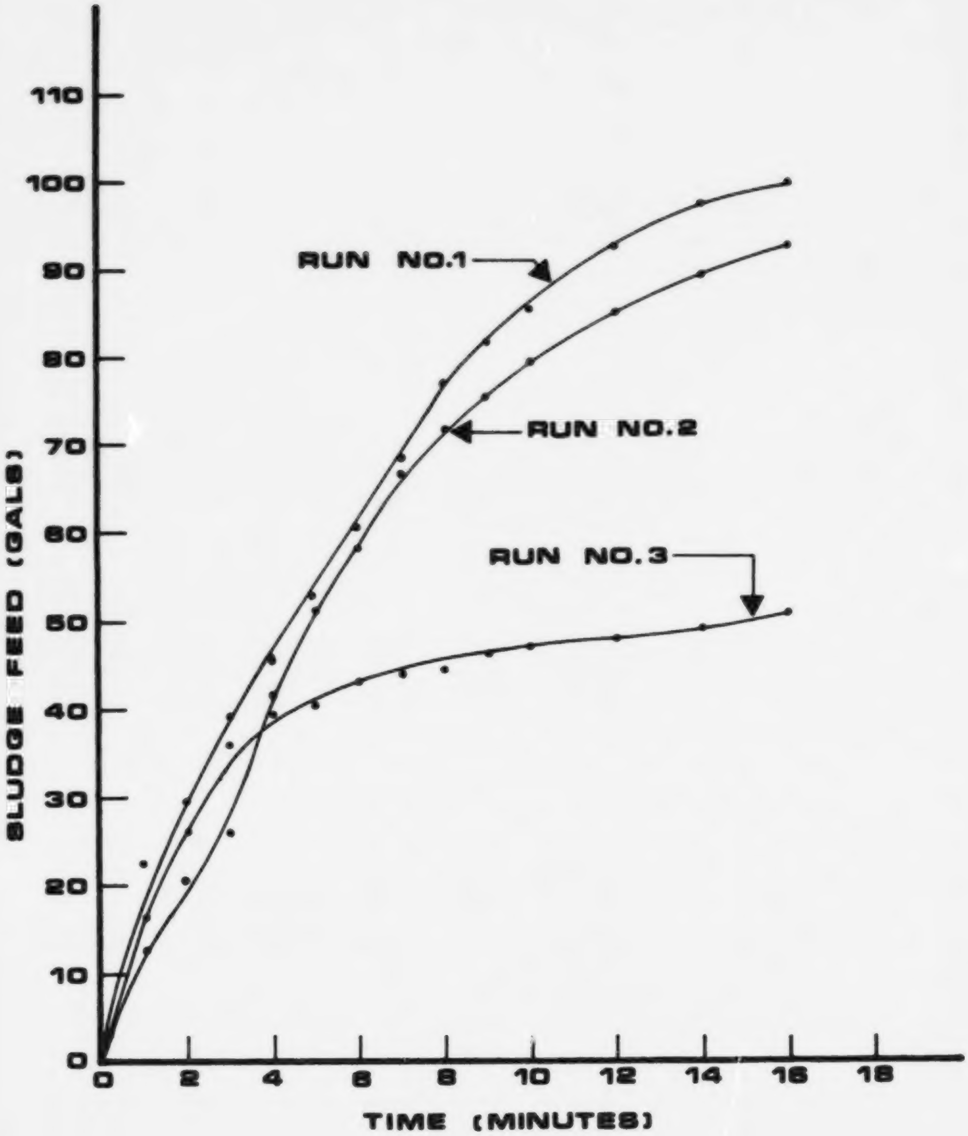


Figure 11. Feed Volume vs. time. NGK runs on 3/4/77.

TABLE 6. RUNS TO OPTIMIZE PUMPING TIME

DATE	% CHEMICALS LIME/$FeCl_3$	CYCLE TIME (MIN) PUMP/SQUEEZE	% CAKE SOLIDS	CAKE DRY WEIGHT (kg)	FULL-SCALE YIELD kg/m^2/hr	REMARKS
4-1-77	15.4/4.5	5/15	44.9	8.0	1.7	thin cake
	15.1/4.3	10/15	42.7	10.9	2.2	thin cake
	15.1/4.6	15/15	33.8	12.1	2.1	
	14.1/4.3	20/15	34.5	13.8	2.2	
4-6-77	18.0/5.4	5/15	43.0	8.0	1.7	thin cake
	18.0/5.4	10/15	39.1	9.4	1.8	thin cake
	18.0/5.4	15/15	37.5	10.7	1.8	
5-19-77	20.4/6.8	19/12	33.6	16.8	2.7	squeeze time too short
	18.6/6.2	15/15	38.0	16.4	2.8	
	13.8/4.6	13/26	35.5	11.3	1.7	marginal conditioning

installation size NGK press. For the 5.8 m^2 pilot press this rate is

$$\frac{2.4 \text{ kg solids/hr/m}^2 \times 5.8 \text{ m}^2}{60 \text{ min/hr}} = .23 \text{ kg solids/min } (.51 \text{ lb/min})$$

For a 7.5% total solids feed and a specific gravity of 1.02, the volumetric feed rate is

$$\frac{.23 \text{ kg solids/min} \times \frac{100 \text{ kg total feed}}{7.5 \text{ kg solids}}}{1.02 \times 1 \text{ kg/l}} = 3.0 \text{ l/min feed } (.8 \text{ gpm})$$

This terminal pumping rate, through calibration of the NGK mix tank, was found to be equivalent to 1/8 inch per minute. During later test work (i.e. after 4/28/77) when the sludge flow rate dropped to 1/8 inch per minute for three consecutive minutes, the pumping cycle was terminated. The variable output of the sludge feed pump and the difficulties in measuring 1/8 inch necessitated a three minute time to ensure that a good measurement was taken.

The runs on 5/19/77 (Table 6) show the effects of this procedure on three different levels of sludge conditioning. The first run, with good conditioning, gave a high yield with a rather long pump time of 19 minutes (low cake solids resulted from an error in the determination of the squeezing time). The second run was still with good conditioning and a high yield resulted. The third run, with marginal conditioning, achieved the required pumping rate in only 13 minutes. This indicated to the operators that the sludge was not well conditioned. It was then necessary to squeeze for a slightly longer time, so that a good cake release would result.

A secondary advantage of using this procedure involved the response of the press to poorly conditioned sludges. In general, with a poorly conditioned sludge, this rate was usually achieved in 5 to 10 minutes and a thin cake was produced. This thin cake, however, would further dewater under extended squeezing times and thus give a good discharge from the filter cloth. Earlier runs showed that poorly conditioned sludge, when allowed to form a thick cake, did not dewater well under extended squeezing and cake sticking and resultant cloth blinding occurred. This new operational procedure thus gave a way for the filter press to compensate for errors that had occurred in the conditioning step. This same method, applied to a well conditioned sludge produced a thick cake with maximized yield. In effect, this method gave the best filter performance for the sludge and conditioning available.

Recognizing the problems with instrumentation that could occur in obtaining a sludge flow rate on a larger filter press, we examined two other methods for optimizing the pump time: 1) rate of pump pressure buildup and 2) filtrate flow rate. The rate of feed pump pressure buildup gives an indication of the resistance to filtration that exists during the dewatering process. It can be used to indicate a poorly conditioned sludge and alert the operator to take corrective action. It cannot be used, however, to define the cycle endpoint for a well conditioned sludge. In Figure 12, the feed pressure curves for the runs of 3/4/77 are shown. The best conditioned sludge (Run #1) built up pressure slowly, indicating little resistance to filtration. At the end of the pumping cycle (16 minute mark), the pressure

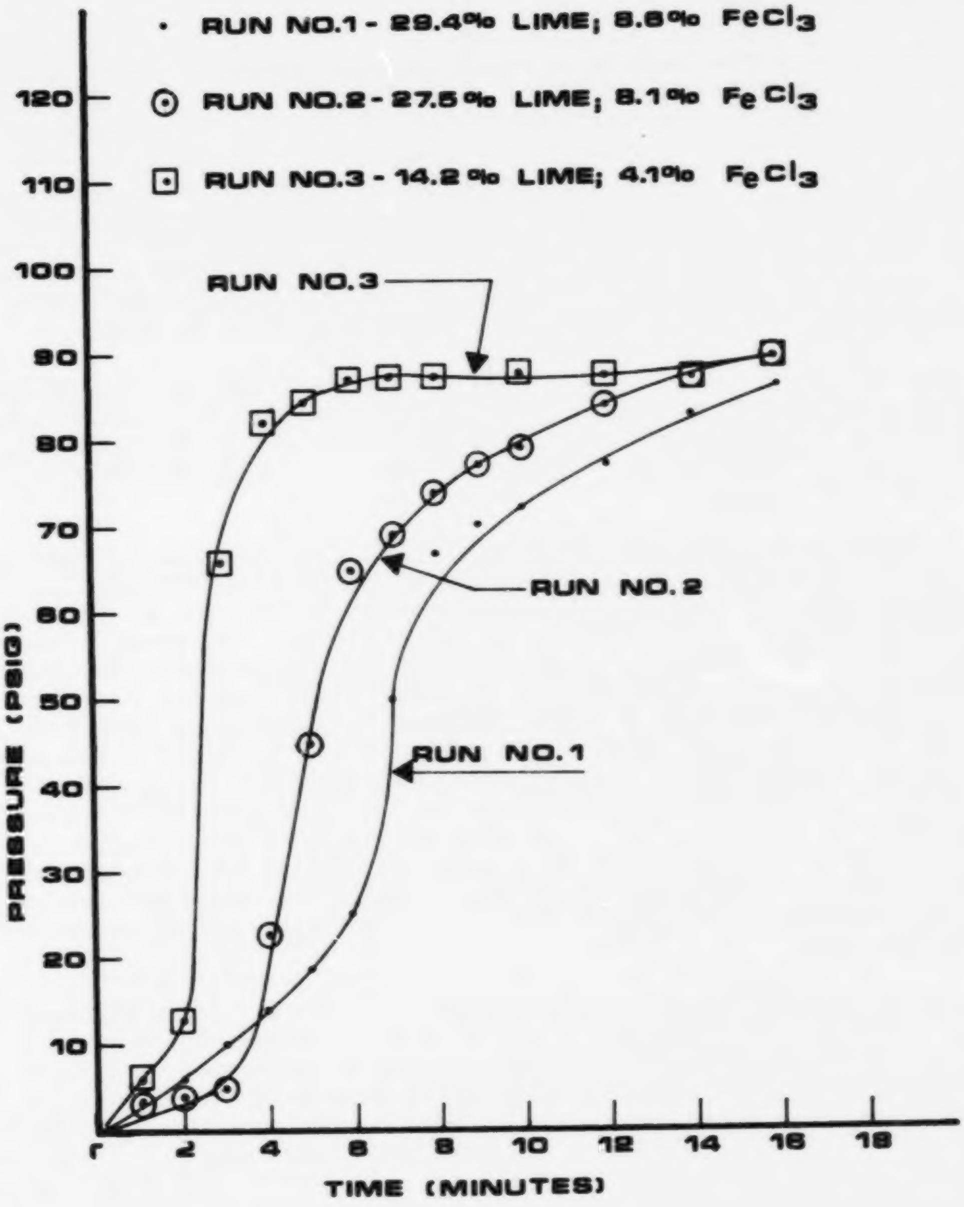


Figure 12. Feed pressure vs. time. NGK runs on 3/4/77.

was still rising at a steady rate. The under-conditioned sludge (Run #3) built up pressure very rapidly. Resistance to filtration was high and the pressure reached a limiting value well before the end of the cycle (also 16 minute mark).

The filtrate flow-rate is an easily measured parameter and a correlation which would determine the end of the pumping cycle can be readily developed from the filtrate flow rate data collected. Figure 13, for example, shows the filtrate collected vs time for the 3/4/77 runs. At the 16 minute mark (the end of the pump cycle), Runs #1 and #2 still had a fairly steep slope, whereas Run #3 had already begun to flatten out. These curves show that since filtrate for run #1 and #2 was still being discharged at a high rate at the end of the cycle, had the pumping times for these runs been extended higher yields would have resulted. Run #3 pumping, though, should have been terminated near the 8 minute mark, where the slope began to flatten out.

Because of the need to test different ratios of blended sludges and variable terminal pump pressures, we decided to use the more direct method of measuring the sludge feed rate to determine the end of the pumping cycle. On a full-scale installation, however, either sludge feed rate or filtrate flow rate could be used.

Squeezing pressure--The squeezing pressure was developed by applying pressurized water to the diaphragms. During startup, the NGK engineers recommended that the press be operated at 15 kg/cm^2 (213 psig). However, the pump could deliver any pressure up to 17.6 kg/cm^2 (250 psig). Numerous tests were run throughout the study period to try to optimize the squeezing pressure level and the rate at which it was applied. None of the tests showed a marked difference in final cake solids or squeezing time over the range of 7 to 17.6 kg/cm^2 (100 to 250 psig). The Blue Plains sludge appeared to dewater independently of pressure within this range. Under normal operation, the full pressure was applied to the diaphragm immediately after the sludge pump stopped. Tests were also run applying the pressure in step increments up to the final squeezing pressure. Again, no difference in results could be determined. Therefore, in nearly all the press runs a squeezing pressure of 15 kg/cm^2 (213 psig), applied as per the manufacturer's design, was used.

Squeezing time--In an optimum pumping cycle for a well conditioned sludge in the NGK press, approximately 75-85% of the filtrate will be collected during pumping. The squeezing cycle is then really a cake consolidation step removing relatively small quantities of filtrate. Generally, the squeezing cycle increases the cake solids from approximately 20% at the end of pumping to 35-40% solids before discharging. Figure 14, for example, shows the results of tests run on 4/13/77. For these five runs the lime/$FeCl_3$ dosage was 20.7%/6.2%; the pump time was 18 minutes. As the squeeze time was increased from 5 to 25 minutes, the cake solids increased from 25% to over 40%, with a corresponding decrease in process yield.

During the initial part of the study (prior to 4/28/77), the squeezing time was preset by the operator. As with the pumping time, however, this method was good only if the sludge and conditioning remained constant. With

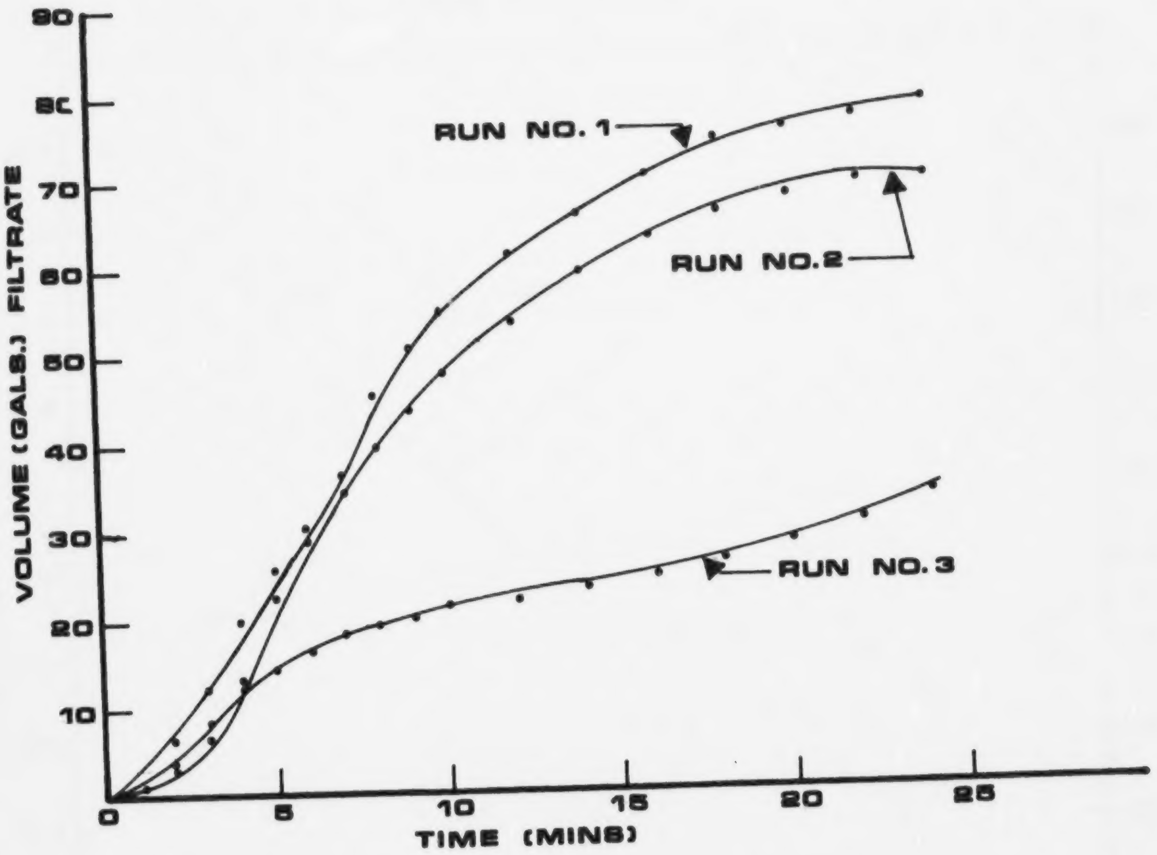


Figure 13. Filtrate volume vs. time. NGK runs on 3/4/77.

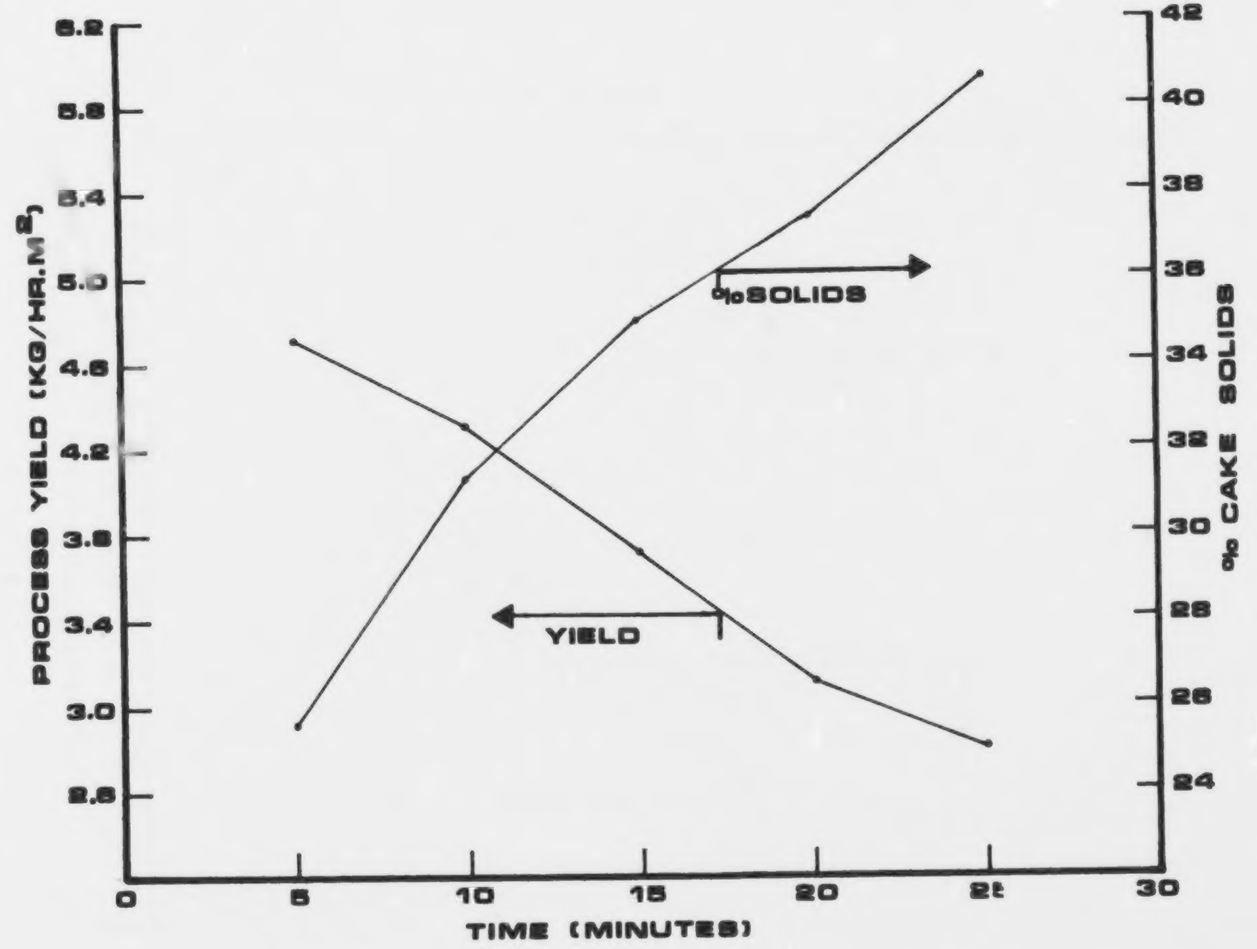


Figure 14. Effect of increasing squeeze times - NGK press.

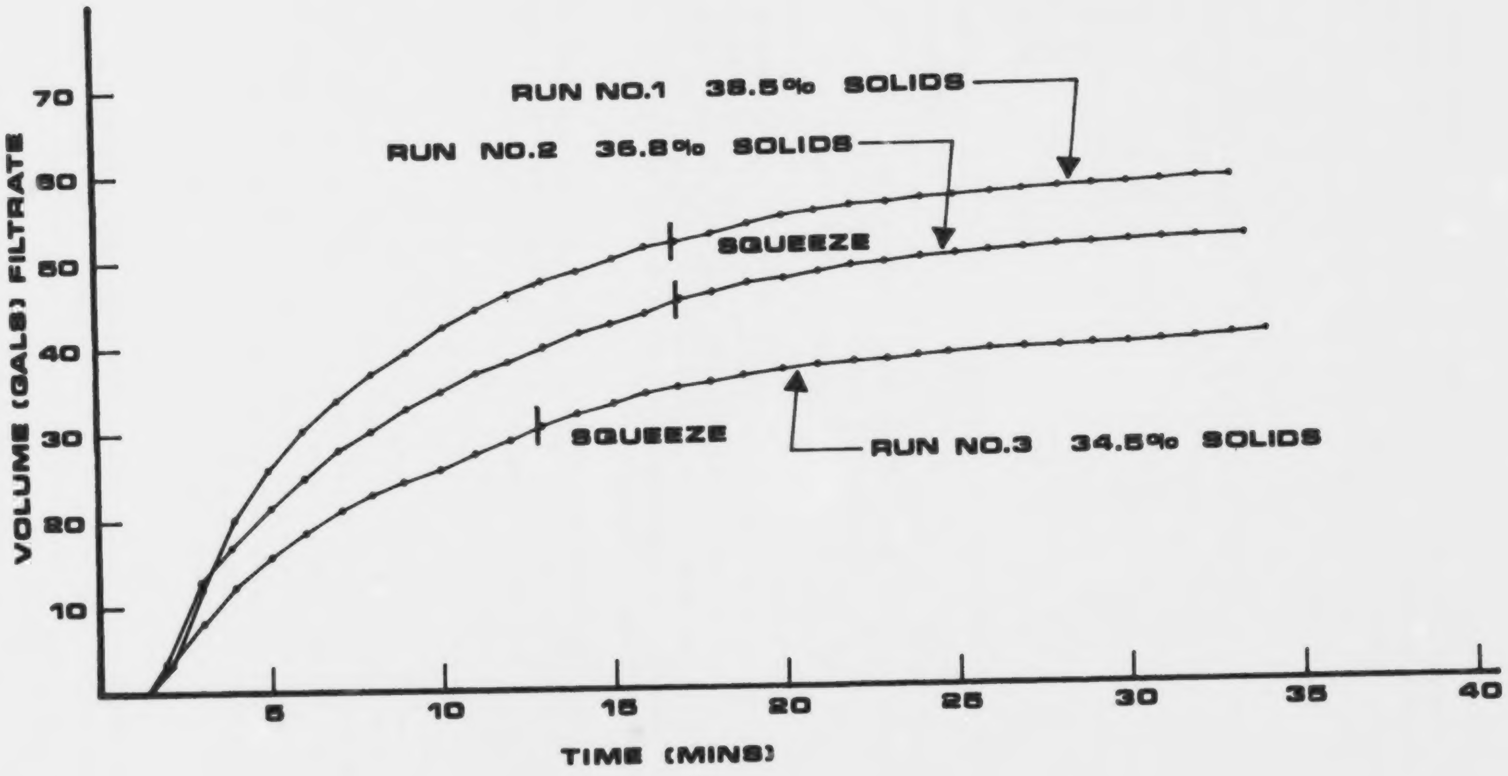


Figure 15. Filtrate volume vs. time. NGK runs on 11/1/77.

a well conditioned sludge, squeezing times of 8 to 10 minutes were normally sufficient to produce the required 35% solids cake; with the marginally conditioned sludges squeezing times of over 20 minutes were required to reach this level. Examination of many press runs showed a correlation between the terminal filtrate flow rate and final cake solids. The data from the runs on 4/13/77 is typical:

Squeeze Time (min)	% cake solids	Terminal filtrate rate l/min	(gpm)
5	25.5	3.78	(1.00)
10	31.3	1.89	(0.50)
15	35.0	0.87	(0.23)
20	37.3	0.57	(0.15)
25	40.6	0.00	(0.00)

Based on results such as these, a procedure was developed to ensure final cake solids of 35% for each run. Beginning on 4/28/77, after pumping the sludge feed to a rate of 3.0 l/min, the squeezing cycle was extended until a terminal filtrate rate of 0.57 l/min (0.15 gpm) was obtained. When this rate held steady for three consecutive minutes (to obtain good measurement), the squeezing cycle was terminated. Subsequent tests showed that this terminal rate method worked well with all levels of conditioned sludge. Figure 15 shows the filtrate curves on three levels of conditioning 100% secondary sludge for this procedure. Data for these 11/1/77 runs are included in Table 5. Run #1 was slightly over-conditioned; Run #2 was conditioned at an average level, and Run #3 was marginally conditioned. In each case the press cycle was the optimum for the type of sludge dewatered; cake solids of about 35% resulted for each run.

Filtrate quality--The overall quality of the filtrate is affected not only by the selection of cloth, but also by the chemical conditioning. Normally, only total solids and suspended solids were analyzed on filtrate samples. On 3/8/77, tests were run to characterize the filtrate for other parameters. The filter cloth was not washed between these eight runs, so that a normal plant operation would be simulated. Feed solids were 6.0% on a 1/1 ratio of secondary/primary sludge. Cycle time for each of the runs was 16 minutes pumping and 8 minutes squeezing. Table 7 shows the filtrate parameters analyzed. Because of filtration difficulties at that time, all runs were made with high chemical dosages. Each of the pollutant levels is consistent with that for a water in contact with undigested sludge. The average percent total volatile solids in the filtrate was 24.8%, indicating a high soluble chemical content (mostly lime). The average percent volatile suspended solids was 60%, indicating that filtrate suspended solids were mostly organic.

Further tests were conducted to determine the effect that chemical conditioning had on filtrate quality. See Table 8. The first three runs on 3/10/77 were well conditioned and gave good cake results. Note that the filtrate total solids (mostly lime) decreased as the chemical addition rate decreased. The last run was underconditioned and gave poor cake results and

TABLE 7. EXTENDED RUNS - 3/8/77

% CHEMICALS LIME/$FeCl_3$	pH	FILTRATE mg/l								
		BOD_5	COD	PO_4	TKN	NH_3	NO_3	TOTAL ALKALINITY	TOTAL SOLIDS	SUSPENDED SOLIDS
26.7/7.8	11.9	554	1560	35.8	173	85.3	.70	1801	7266	144
26.8/7.8	11.9	916	2067	110	253	107	.74	2367	9431	1540
26.8/7.8	11.9	694	2061	63.6	214	105	.80	1987	8575	464
29.3/8.5	11.7	684	1640	44.7	244	106	.75	2021	8307	151
29.3/8.5	11.7	682	1668	56.7	240	96.9	.77	1961	8203	62
29.3/8.5	11.8	684	1785	69.2	296	99.2	.79	1959	8140	55
30.0/8.8	11.8	420	1730	47.5	226	96.9	.74	2063	8471	127
30.0/8.8	11.9	374	1867	64.0	253	110	.72	2008	8474	47
28.5/8.3	11.8	626	1923	61.4	238	101	.75	2020	8361	323 (Average)

TABLE 8. FILTRATE QUALITY VS. CHEMICAL CONDITIONING

DATE	% CHEMICALS LIME/$FeCl_3$	% CAKE SOLIDS	CAKE DISCHARGE	TOTAL SOLIDS mg/l	SUSPENDED SOLIDS mg/l	pH
3-10-77	45.5/13.3	39.6	excellent	10120	87	11.5
3-10-77	34.4/10.0	38.5	excellent	8910	70	11.6
3-10-77	22.8/6.7	36.0	excellent	8189	80	11.5
3-10-77	17.1/5.0	31.6	good	9404	2604	11.5
7-12-77	16.8/5.6	41.8	excellent	9815/10130*	120/18*	11.5
7-12-77	14.4/4.9	39.9	excellent	9736/10024*	216/45*	11.5
7-12-77	12.2/4.1	29.0	poor	8963/8057*	438/198*	11.5
11-1-77	25.2/8.3	38.5	excellent	6773	28	-
11-1-77	19.7/6.6	36.8	excellent	5879	22	-
11-1-77	14.8/5.0	34.5	good	5546	189	-

* First entry is from the pump cycle; second entry from the squeezing cycle.

high filtrate suspended solids. High filtrate suspended solids were common for a poorly conditioned sludge. The fine particles were not well flocculated and passed through the filter cloth. The runs on 11/1/77 show identical results.

The following typical run shows the filtrate solids as sampled at various time intervals during the pumping cycle. This was taken from Run #1 on 5/3/77 shown in Table 5.

Time (min)	Filtrate Total solids mg/l	Filtrate Suspended solids mg/l
1.5	8242	716
3.0	8169	550
5.0	7687	55
10.0	7655	45
18.0	7604	19

The suspended solids drop off rapidly after three minutes. This is due to the initial filtration being through the cloth. Once a cake is formed in the chambers, the accumulated solids then act as the primary filter media. The composite sample of filtrate during the pumping cycle averaged 7826 mg/l total solids and 308 mg/l suspended solids. The squeezing cycle for the same run showed 7713 mg/l total solids and 50 mg/l suspended solids. Note that the suspended solids were lower for the squeezing cycle than for the pumping cycle. This was true for all the press runs. The runs on 7/12/77 (Table 8) show several cases in which the filtrate collected from the pumping and squeezing cycles were analyzed separately.

Filter cloth evaluation--Filter cloth selection depends on resistance to wear and abrasion, the ease of cake release, and filtrate quality desired. Three different filter cloths were supplied with the NGK filter press, each of which was tested during the study. Because of the limited number of runs made and the method of operation (day time only), resistance to wear and abrasion could not be determined by this study. Cloth media specifications are given in Table 9.

TABLE 9. NGK FILTER CLOTHS

TYPE	CONSTRUCTION	MATERIAL WARP/FILLING	AIR PERMEABILITY at Δ P = 12.7 mm H_2O $Cm^3/Sec/Cm^2$
NY 516	Plain	polyamide/polypropylene	4.0
TR 520	Herringbone twill	polyester/polyester	11.0
NY 51-4	Twill	polyamide/polyester	93.0

The NY 516 cloth, the tightest weave, was the first cloth tested. A total of 151 runs (5/26/76 thru 11/1/76) were made on this cloth, mostly with plant thickened sludge. Filtrate analyses for 71 of these runs showed excellent results with an average of 16.7 mg/l suspended solids. Cake discharge, however, was not always the best. For nearly all the runs, except those that were over conditioned, the cloth shaker was needed for cake discharge. The cloth surface was rough and particles of cake tended to stick. Underconditioned sludges blinded the cloth very readily and excessive scrubbing was required. The cloth washing system, operated at 350 psig, did little to rejuvenate the cloth after such a run. Resistance to abrasion and wear seemed to be very high.

The second type cloth tested was the NY 51-4 media. This cloth had the most open weave and a very smooth surface. With the exception of very poorly conditioned sludges, cake discharge was almost always good. The first set of NY 51-4 cloths, operated from 11/2/75 through 4/7/77, were worn badly after 182 runs due to over-zealous brushing while cleaning. The brushing was required during early 1977 when difficulties were encountered with dewatering the sludge. The second set of NY 51-4 cloths were operated for 213 runs from 4/8/77 to 10/7/77. Filtrate analyses on 271 samples with both NY 51-4 cloths averaged 525 mg/l suspended solids on all types of sludge. This cloth gave the best overall cake solids, since it provided little resistance to filtration. Filtrate quality was, however, a drawback. The cloth also seemed to show little resistance to wear. The second set were beginning to tear in places after only 130 runs, but this may have been due to the alternate wetting and drying (which is known to stretch fibers) caused by our operational schedule. Further evaluation on a continuous basis is required before any definite conclusions can be made.

The TR 520 cloths were tested from 10/13/77 through 11/30/77 for a total of 80 runs. Only 22 of these runs were analyzed for filtrate quality giving 52.4 mg/l suspended solids. Even though this was a textured, heavy cloth, cake discharge was excellent--equal to the NY 51-4 cloth. Sufficient runs to determine abrasion resistance were not made, but the cloth seemed to be more sturdy than the NY 51-4 cloth. The TR 520 cloth appears to provide the best compromise for both good discharge and acceptable filtrate quality.

Cloth washing requirements are more a function of chemical conditioning than cloth selection. When the conditioning was optimum, up to 15 runs were made on each of the cloths before they required washing. The cloth washing system, while designed to operate at 70 kg/cm^2 (1000 psig), only produced a maximum pressure of 24.6 kg/cm^2 (350 psig). A defective pressure gage and regulator valve caused this problem but was not discovered until all tests were completed.

Filter yield--A major purpose of the study was to develop design parameters to dewater a 2/1 ratio of secondary/primary sludge from an initial 5% solids mixture to a 35% solids cake. A total of 142 runs were made on this sludge ratio. Recognizing that sludge variability was an important factor during the study and that many types of experiments were made, only the runs that gave at least a 35% solids cake were used to produce representative design conditions for dewatering the Blue Plains sludge on a year-round basis. According to the manufacturer, scale-up from the pilot press data to a full-scale unit can be made directly. Filtration (pumping) and squeezing cycle times will be identical. A mechanical turn-around time, though, must be included in the total cycle time in order to obtain the full-scale yield. (See Appendix B, Explanation of Data Sheet 4 for details).

These runs, tabulated in Table 10, are summarized below:

Chemical dosage	19.6% lime/6.5% $FeCl_3$
Cycle time	16.9 min pumping/ 18.1 min squeezing
Final cake solids	38.7%
Full-scale yield	2.39 $kg/hr/m^2$ (0.49 $lb/hr/ft^2$)

This data was used to develop average design parameters for a full-scale press installation at Blue Plains. Cost estimates for the NGK press, given in Section 9, are based on these average values.

Special Tests--

Dewatering of variable sludge ratios--A secondary purpose of the study was to observe the effect of dewatering various ratios of secondary to primary sludge solids. Tests on the Buchner funnel, Figure 3, had shown that more chemicals were required as the percentage of secondary sludge increased. Tests on the NGK press confirmed these results and also showed the effects that the sludge ratio had on filter yield. During the month of August, 1977 the sludges were fairly consistent in their filterability. During that month, seven different sludge ratios were tested. With each sludge ratio at least three runs at three different chemical dosages were made; one over-conditioned, one average conditioned, and one marginally conditioned. The results are averaged for each sludge ratio in Table 11. Note the general trend that sludges high in primary solids give high cake solids and high yields with relatively low chemical dosages. Once the ratio of solids increases above 1/1 secondary to primary, the secondary sludge is the controlling factor and the sludges become more difficult to dewater.

Three day continuous run--From 10/4/77 - 10/7/77 the NGK pilot unit was operated continuously for a period of 72 hours. The primary objectives of this test were:

a. To simulate a full-scale installation and thereby obtain representative operating parameters,

b. to test the effectiveness of a continuous chemical conditioning scheme;

c. to establish diagnostic and monitoring procedures for a full-scale system,

d. to operate the unit under stress conditions in order to evaluate the mechanical design, and

TABLE 10. NGK RUNS ON 2/1 SLUDGE

DATE	% CHEMICALS LIME/$FeCl_3$	CYCLE TIME (MIN) PUMP/SQUEEZE	% CAKE SOLIDS	FULL-SCALE Yield $kg/hr/m^2$
3-15-77	26.0/8.1	30/15	41.7	2.78
	16.0/4.6	30/15	44.4	3.30
3-16-77	15.9/4.6	30/15	36.4	2.54
	19.6/5.7	20/15	39.9	2.94
3-30-77	21.6/6.3	20/15	41.4	3.23
	21.6/6.3	20/15	40.6	3.13
3-31-77	14.9/4.2	20/15	38.8	2.72
	15.0/4.2	20/15	35.8	2.69
4-1-77	15.4/4.5	5/15	44.9	1.75
	15.1/4.3	10/16	42.7	2.17
4-5-77	21.9/6.4	5/15	43.8	1.93
	21.9/6.4	10/15	38.3	2.27
4-7-77	20.8/6.0	5/15	41.3	1.78
	20.8/6.0	10/15	40.0	2.47
	20.3/6.0	15/15	38.8	2.75
	20.3/6.0	20/15	36.7	2.28
	20.3/6.0	25/15	37.7	2.52
4-12-77	17.2/5.0	18/25	38.3	1.89
4-13-77	20.7/6.2	18/15	35.0	2.32
	20.7/6.3	18/20	37.3	2.05
	21.6/6.6	18/25	40.6	1.93
4-26-77	21.5/6.4	20/20	36.6	2.09
4-28-77	19.3/5.9	17/20	37.8	1.78
	24.0/6.9	22/20	39.7	2.18
4-29-77	19.5/6.5	20/19	35.7	2.07
	25.5/8.4	21/19	39.8	2.35
	15.5/5.2	18/25	35.1	1.66
5-3-77	18.6/6.2	18/19	38.0	1.97
	17.4/5.8	9/24	38.1	1.30
	13.7/4.5	6/20	37.5	1.05
5-4-77	24.4/8.4	19/14	37.1	2.57
	20.2/6.8	16/16	37.8	2.48
5-11-77	19.4/6.2	15/22	36.6	1.65
5-12-77	19.6/6.4	17/22	38.0	1.87
	20.2/6.6	15/21	37.0	1.82
5-17-77	17.6/5.9	11/23	37.7	1.63
5-18-77	20.3/6.8	18/19	35.8	1.87
5-19-77	18.6/6.2	15/15	38.0	2.78
	13.8/4.6	13/26	35.5	1.71
5-20-77	19.9/6.8	11/22	35.8	1.46
5-23-77	17.1/5.8	16/18	36.1	1.91
	19.5/6.5	19/20	38.1	2.17
5-24-77	14.6/4.9	18/16	37.4	2.48
5-26-77	20.3/6.6	18/19	36.1	2.00

TABLE 10.

DATE	% CHEMICALS LIME/$FeCl_3$	CYCLE TIME (MIN) PUMP/SQUEEZE	% CAKE SOLIDS	FULL-SCALE Yield $kg/m^2/hr$
6-15-77	19.9/6.6	16/15	36.5	2.21
6-22-77	20.4/6.9	11/21	41.7	1.96
	25.4/8.4	17/18	35.5	1.93
6-23-77	24.7/8.3	15/18	39.2	1.95
	27.0/9.0	16/18	38.4	2.07
7-8-77	20.6/6.8	16/16	41.0	3.22
7-11-77	25.9/8.7	17/16	39.6	2.60
	13.9/4.6	18/20	37.7	2.64
7-12-77	19.0/6.4	14/17	39.2	2.54
7-14-77	20.9/7.0	20/22	42.5	2.22
7-27-77	26.8/8.9	16/14	44.9	3.70
7-28-77	16.8/5.6	17/19	40.7	2.97
8-2-77	22.1/7.4	21/16	41.4	3.07
	17.9/5.9	20/18	40.8	2.83
8-4-77	14.7/4.9	14/23	40.2	2.16
8-17-77	18.0/6.0	16/15	39.9	2.91
	17.5/5.8	16/16	39.1	2.70
8-18-77	19.6/6.6	18/16	40.2	2.76
8-19-77	21.5/7.2	20/16	42.2	3.19
	14.2/4.8	18/21	36.0	2.10
8-23-77	16.6/5.5	15/20	36.4	2.12
8-24-77	18.3/6.2	21/23	41.4	2.19
8-25-77	20.1/6.7	18/18	43.0	3.07
	19.3/6.5	17/17	40.9	2.76
9-1-77	17.9/6.0	20/20	38.8	2.35
9-2-77	18.9/6.3	15/18	38.7	2.45
	17.8/6.0	16/21	38.5	2.39
9-9-77	19.1/6.4	16/22	37.2	2.05
	19.1/6.4	12/22	36.7	1.88
9-14-77	24.3/8.1	13/22	36.3	1.90
	29.2/9.7	19/17	35.5	2.30
9-21-77	27.7/9.2	16/16	41.0	2.67
9-22-77	24.1/8.0	18/20	36.1	1.79
10-13-77	15.0/5.0	20/15	35.7	2.69
10-18-77	23.4/7.8	18/17	37.3	2.44
	20.1/6.6	14/15	41.0	3.18
	14.8/5.0	17/18	35.5	2.47
10-19-77	20.0/6.7	18/17	39.2	2.53
	20.0/6.7	17/19	38.0	2.33
	20.0/6.7	15/17	45.3	3.03
	20.2/6.8	18/17	37.1	2.66
	20.2/6.8	19/17	38.3	2.59
	20.9/7.0	17/18	34.7	2.25

TABLE 10.

DATE	% CHEMICALS LIME/$FeCl_3$	CYCLE TIME (MIN) PUMP/SQUEEZE	% CAKE SOLIDS	FULL-SCALE Yield $kg/m^2/hr$
10-25-77	19.3/6.4	19/15	37.1	2.82
10-26-77	25.7/8.5	19/14	41.7	2.42
	15.8/5.3	18/18	36.9	2.62
	15.8/5.3	16/18	37.1	2.63
10-27-77	12.0/4.0	18/20	35.5	2.31
	12.1/4.0	15/21	36.2	2.05
10-28-77	19.7/6.6	20/16	40.5	2.94
11-2-77	20.4/6.8	20/17	39.8	2.62
Averages	19.6/6.5	16.9/18.1	38.7	2.39

TABLE 11. TYPICAL RESULTS ON DIAPHRAGM PRESS - AUGUST RUNS

SEC/PRIM RATIO	NO. OF RUNS	% CHEMICALS LIME/$FeCl_3$	PROCESS CYCLE/TIME (MIN)	% CAKE SOLIDS	FULL-SCALE YIELD ($kg/hr/m^2$)	FILTRATE SOLIDS mg/1 TOTAL/SUSPENDED
0/1	2	13.4/4.5	23	54.0	4.95	8366/1926
1/2	2	13.2/4.4	32	48.1	3.62	8156/602
1/1	3	15.6/5.2	35	47.4	3.46	8558/308
2/1	13	18.3/6.1	36	40.0	2.63	8238/386
3/1	5	22.7/7.6	39	40.7	2.41	10358/380
4/1	3	21.4/7.2	34	43.1	3.06	7927/197
1/0	4	23.5/7.8	34	38.9	2.28	7971/297

e. to acquaint plant engineers, maintenance, and operating personnel with the design and operation of a filter press.

The flowsheet for the process is shown in Figure 16. Primary sludge (1% solids) and secondary sludge (0.75% solids) were gravity thickened to 8% and 4% solids, respectively, and mixed in the blending tank in a secondary/primary solids ratio of 2/1. A recycle rate within this tank of 10-20 gpm provided the necessary agitation for mixing. From here, the blended sludge (5.3% solids) was pumped continuously at 1.5 gpm to the chemical conditioning system. A Komline-Sanderson Rotary Drum Conditioner, with internal baffles for mixing, was used as the conditioning tank. Ferric chloride (13% weight solution) was added by a positive displacement pump to the sludge feed line; lime slurry (6.5% - 13% weight solution) was added to the middle of the conditioning drum. In order to minimize floc deterioration during the conditioning step, drum speed was maintained at one RPM, and the sludge was detained only a few minutes before overflowing to the press feed tank. This tank held enough sludge for 2-3 press runs and had an average sludge retention time of 1.5 hours.

Automatic filtration and squeezing cycles of about 20 minutes each were used during press runs. These cycles were checked and adjusted every fifth run by measuring the sludge and filtrate flow rates. The cloth washing cycle was initiated only when required; hence, the turn-around time between successive runs averaged only 10 minutes.

Sludge filterability was monitored each run by Buchner funnel and CST tests on the sludge leaving the conditioning tank. A CST of 15 seconds and a Buchner funnel filtrate rate of 80 ml/2 min was used as an indicator of acceptable filterability. Samples of sludge, cake, and filtrate were taken every fifth run for laboratory analysis.

A total of 76 runs were made, 72 on the 2/1 secondary/primary mixture and 4 on the 100% secondary sludge. Approximately 5600 gallons of sludge were filtered during the operation. Press down-time was minimal and 59.3 operating hours were logged. Results of the laboratory analyses are summarized in the following table:

Sludge feed solids/volatile solids	5.32%/67.5%
Conditioned sludge volatile solids	46.9%
Lime dosage (average)	22.9% (of sludge solids)
$FeCl_3$ dosage (average)	7.3% (of sludge solids)
Cycle time (avg) pump/squeeze/ mechanical	20/20/5 minutes
Cake wet weight (total)	2682 kg
Cake solids/volatile solids	36.3/48.2%
Cake dry weight	973 kg
Cake sludge solids	748 kg
Yield (average)	2.17 $kg/hr/m^2$
Filtrate suspended solids	83.2 mg/l
Filtrate total solids	9465 mg/l
Cloth washed	16 times (every 4.75 runs)
Cloth used	NY 51-4

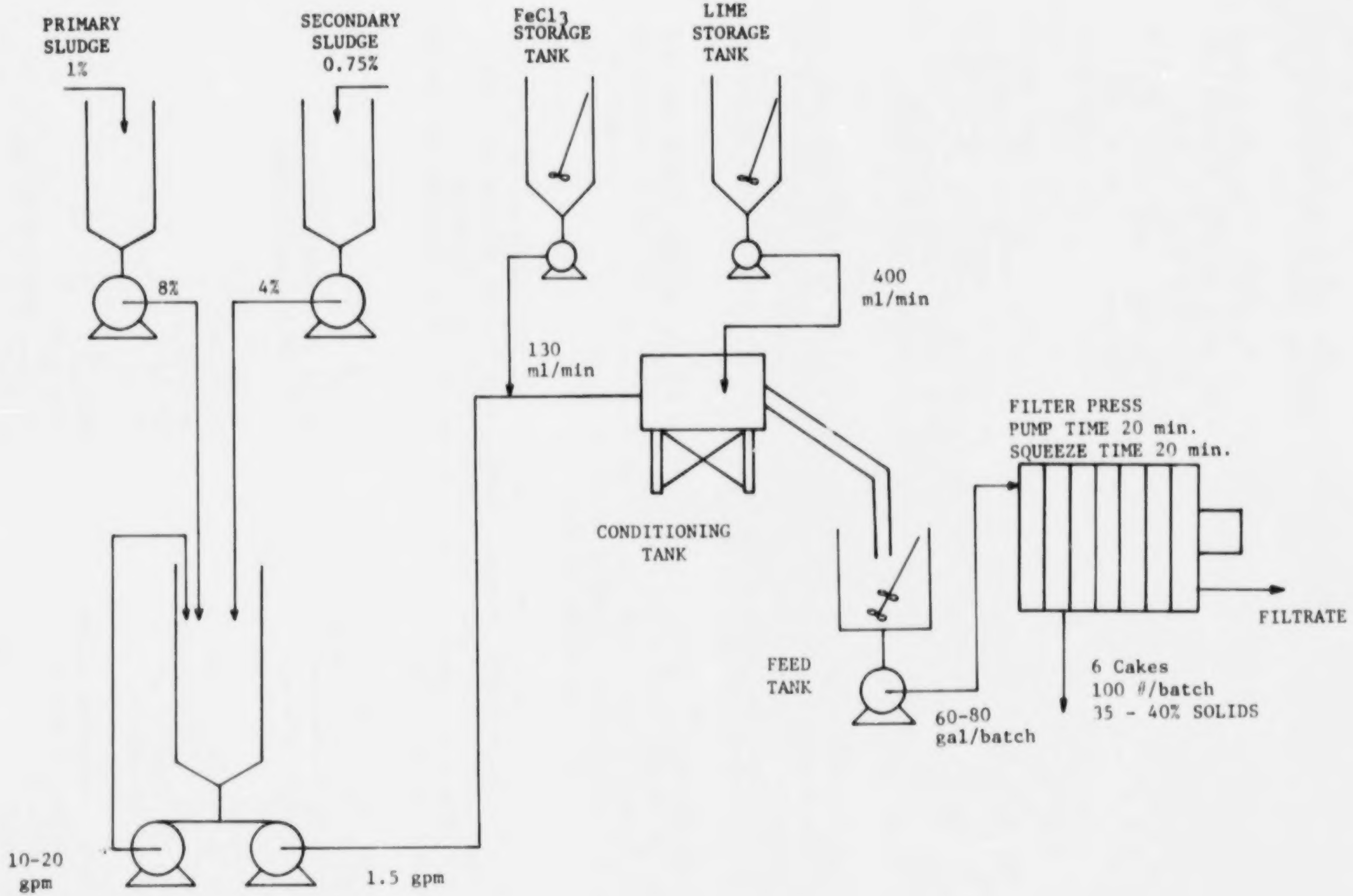


Figure 16. Process flowsheet for continuous run.

The chemical addition system proved to be the major bottleneck of the operation. Frequent plugging and clogging of the lime slurry feed line caused the sludge feed to be under-conditioned for several of the runs. Consequently, poor cake discharge and excessive cloth blinding occurred. Because of these problems in the lime system, the bench-scale filterability tests were invaluable. Interruptions in the lime delivery were immediately evident by high CST values (e.g. 257 sec) and low Buchner funnel filtrate rates (e.g. 14 ml/2 min). Thus, the operators were able to slug dose the feed tank in order to avoid disastrous press runs.

Problems with the electrical functions of the cake discharge and cloth wash mechanisms caused some minor delays in the automatic operation of the press. But otherwise, the press performed extremely well during the extended operation.

Plant personnel, who initially were unfamiliar with the press, were generally pleased with its operation.

The entire project, therefore, was considered highly successful. Several important design suggestions evolved from these continuous runs and will be discussed later in the design section of the report.

Lasta Diaphragm Press

From 10/25/77 to 11/3/77 Ingersoll-Rand, Nashua, New Hampshire, provided a trailer mounted demonstration unit of their Lasta press for testing. This unit was also a Japanese-made press and is manufactured under license from Ishigaki Mechanical Industry Co., Ltd. Tests were run for comparison with the NGK diaphragm press.

Facilities--

1. Press - 1.64 m^2 (17.65 ft^2) filtration area; contained four chambers with eight 600 mm (23.6 inches) square plates; every other plate was equipped with concave rubber diaphragms. See Figure 17.
2. Tank assembly
    a. sludge conditioning tank - 0.7 m^3 (184 gal) tank with variable speed mixer
    b. sludge storage tank - 1.3 m^3 (350 gal) tank
    c. lime slurry tank - 1.3 m^3 (350 gal) tank, with constant speed mixer
    d. ferric chloride tank - 0.6 m^3 (150 gal) tank, with constant speed mixer
    e. water storage tank - 0.6 m^3 (150 gal) tank

3. Pump assembly
    a. three (3) constant speed pumps - cake wash, cloth wash, diaphragm pressurization
    b. three (3) variable speed pumps - sludge, lime, and ferric chloride delivery
    c. one (1) vacuum pump - diaphragm deflation
    d. hydraulic pump - opening and closing press
4. Air compressor with receiver - core blow and instrumentation control
5. Filter cake conveyor

Operation--

Conditioned sludge was prepared in the NGK mix tank and pumped to the Lasta conditioning tank for use during the tests. As with the NGK press, the Lasta press cycle included pumping, squeezing, cake discharging, and cloth washing operations. The pumping cycle, during which sludge was fed to the press, averaged 10 minutes at pressures of 4.6 - 7.0 kg/cm^2 (65-100 psig). Sludge feed volume ranged from 4.5-60.9 liters (1.2-16.1 gallons) and entered the filtering chambers via special dispersion nozzles located at the top

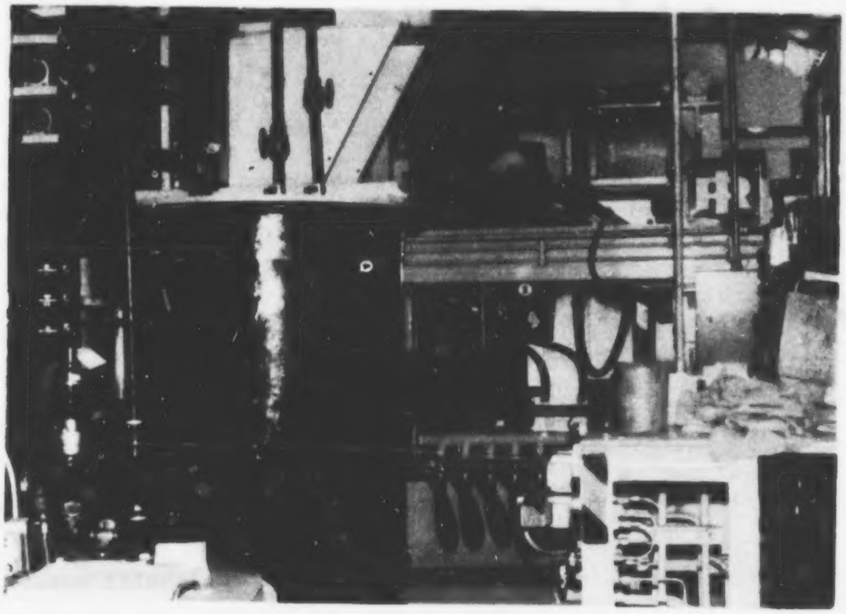


Figure 17. Lasta Diaphragm Press.

center of the filter plates. Filtrate was discharged through side nozzles at the bottom of the filter plates.

At the termination of the pumping cycle, the squeezing cycle began immediately. An average cycle length was 20 minutes at pressures of 14.8 kg/cm^2 (210 psig) provided by stored, recirculated water. At the end of the cycle, the sludge and filtrate lines were blown out by compressed air and the diaphragms were gravity drained and returned to their original shape by vacuum suction.

Cake discharging and cloth washing operations began at the conclusion of the squeezing cycle. As shown in Figure 18, these operations were completely different from those of the NGK press. The Lasta unit released all four cakes simultaneously by a "traveling" motion of the filter cloths. The cloths moved downward around the bottom of the plates in a U-turn fashion which caused the cakes to release. After the discharging was completed, the cloths then returned to their original positions. (The press also had doctor blades located at the bottom of the plates to assist in difficult cake releases.) The filter cloths moved downward a second time for washing on both sides by low pressure, 7 kg/cm^2 (100 psig), spray showers located near the bottom of the press. Drip pans were closed over the discharge port in order to catch spent wash water and prevent rewetting of the filter cake.

Standard laboratory analyses were performed on samples of the sludge, cake, and filtrate.

Test data--

Over the two week test period, 35 runs were made on this press; the results are presented in Table 12. For each batch of sludge, two to three runs were usually made at varying cycle times to optimize the yields and cake solids. For the 2/1 sludge mixture tested, these runs are clustered together in Table 12; for the 100% secondary sludge tested, typical runs at different conditioning levels are shown. As shown by this data, the press performed quite well and produced cake solids of at least 35% in most cases. At this time, though, the feed solids content was high, and the sludge was easily filtered, even at low chemical dosages. (Yields were calculated by adding a full-scale mechanical time of 10.5 minutes to the process cycle time.)

Three different filter cloths, with the following specifications, were tested on the press:

TYPE	CONSTRUCTION	FILLING	THICKNESS mm	AIR PERMEABILITY cm^3/min/cm^2
891	2 x 2 twill	polypropylene	1.46	1500
920	2 x 2 twill	polypropylene	1.17	800
940	2 x 2 twill	polypropylene	1.02	2400

TABLE 12. LASTA RUNS

DATE	% CHEMICALS LIME/$FeCl_3$	CYCLE TIME (MIN) PUMP/SQUEEZE	% CAKE SOLIDS	CAKE THICKNESS (mm)	FULL SCALE YIELD (kg/hr/m^2)
10-25-77	19.3/6.4	8/8	35.1	12	4.05
10-25-77	19.3/6.4	7/10	33.2	11	3.37
10-26-77	25.7/8.5	15/16.5	34.8	20	3.37
10-26-77	25.7/8.5	10/17	34.6	18	3.56
10-26-77	25.7/8.5	5/15	36.1	15	3.56
10-26-77	15.8/5.3	15/17	35.4	20	3.61
10-26-77	15.8/5.3	10/17	34.8	15	3.03
10-26-77	15.8/5.3	5/15	38.0	12	2.64
10-27-77	12.0/4.0	20/30	38.0	18	2.69
10-27-77	12.0/4.0	15/30	39.7	12	2.15
10-27-77	12.1/4.0	10/25	40.7	10	1.32
10-27-77	12.1/4.0	15/30	38.5	8	1.42
10-27-77	12.1/4.0	5/25	39.9	6	1.42
10-27-77	11.9/4.0	5/18	35.3	9	1.54
10-27-77	11.9/4.0	5/22	33.6	7	1.23
10-28-77	19.7/6.6	5/20	41.9	12	3.12
10-28-77	19.7/6.6	10/25	42.5	15	3.03
10-28-77	15.0/5.0	10/25	40.9	14	2.82
10-28-77	15.0/5.0	15/30	42.5	13	2.61
10-28-77	15.0/5.0	5/20	43.6	9	2.99

(continued)

TABLE 12. CONTINUED

DATE	% CHEMICALS LIME/$FeCl_3$	CYCLE TIME (MIN) PUMP/SQUEEZE	% CAKE SOLIDS	CAKE THICKNESS (mm)	FULL SCALE YIELD (kg/hr/m^2)
11-2-77	20.4/6.8	5/15	39.2	11	3.37
11-2-77	20.4/6.8	5/20	40.5	11	3.03
11-2-77	20.4/6.8	8/20	41.6	-	-
*10-31-77	15.8/5.1	5/20	42.6	10	2.61
*11-1-77	19.7/6.6	3/18	37.3	11	2.99
*11-1-77	25.2/8.3	5/15	35.2	-	3.92

*100 % Secondary Sludge

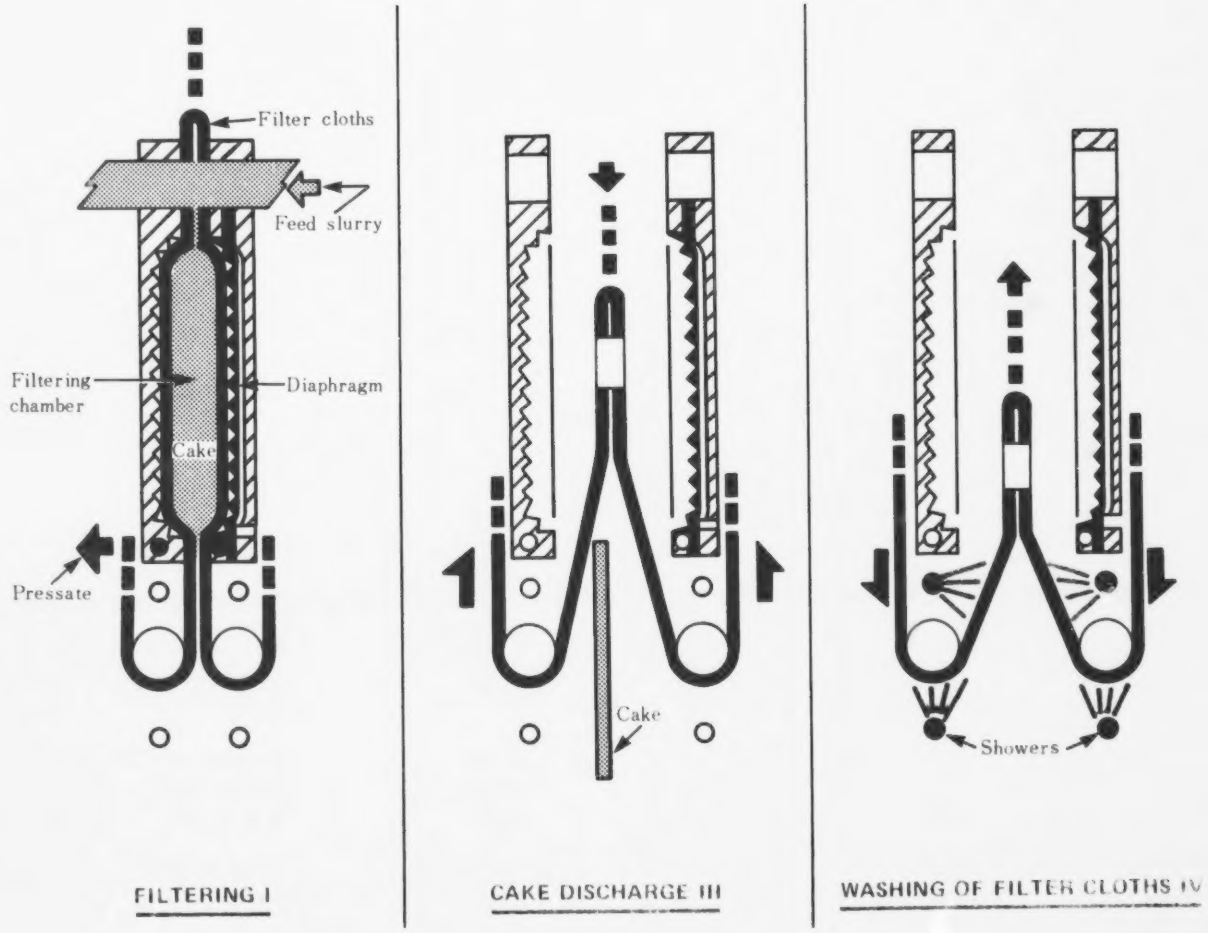


Figure 18. Schematic of filtration, discharge, and washing in the Lasta Press. (Source: Ingersoll-Rand)

The best filtrate quality was obtained using the 891 cloth; suspended solids averaged 87.5 mg/l for runs made on the 2/1 sludge. Cake discharge was good for all three cloths, but sufficient runs were not made to evaluate resistance to wear and abrasion.

Comparison with NGK press--

During the test period, simultaneous runs were made on the NGK press. Sludge was prepared in the NGK mix tank; a portion was pumped to the Lasta conditioning tank and the remainder was fed to the NGK unit.

In Table 13, comparable runs for the 2/1 sludge mixture are shown. The full-scale NGK yield assumes a 19 minute mechanical cycle with cloth washing every 20 runs. The full-scale Lasta yield assumes a 10.5 minute mechanical time with cloth washing every four runs. "Equivalent full-scale yields" were calculated for cycle times at which the lowest cake solids were achieved for either press. The performance of both presses, as shown by the average cake solids achieved, was essentially equal. The main advantage of the Lasta press was its shorter mechanical time (10.5 min vs. 19 min), positive cake discharge, and ease and speed of cloth washing. Additionally, the optimum cycle on the Lasta unit usually had a shorter pump time than the NGK press, which resulted in a thinner cake for discharge.

The main disadvantage of the Lasta unit is the quantity of total filtration area which is currently available on the full-scale Lasta unit. When comparing equivalent yields, the Lasta unit was much higher, averaging 3.31 $kg/hr/m^2$ as compared to 2.70 $kg/hr/m^2$ for the NGK press. This difference represents an additional 22.6% filtration area that the NGK unit would require in order to dewater the same quantity of sludge to the same cake solids as the Lasta press. However, the largest NGK press has 145% more filtration area available than the largest Lasta unit (NGK-500 m^2; Lasta-204 m^2); therefore, fewer NGK units would be required.

TABLE 13. COMPARISON RUNS ON 2/1 SLUDGE

DATE	% CHEMICALS LIME/FeCl	% CAKE SOLIDS		EQUIVALENT % CAKE SOLIDS	EQUIVALENT FULL-SCALE YIELD ($kg/hr/m^2$)		EQUIVALENT CYCLE TIMES (min) PUMP/SQUEEZE		FILTRATE SUSPENDED SOLIDS (mg/1)	
		NGK	LASTA		NGK	LASTA	NGK	LASTA	NGK	LASTA
10-25	19.3/6.4	37.1	35.1	35.0	3.15	4.05	19/15	8/8	26	67
10-26	25.7/8.5	41.7	34.6	34.6	2.93	3.56	19/5	10/17	37	26
10-26	15.8/5.3	36.9	35.4	35.4	2.82	3.61	18/14	15/17	69	43
10-27	12.0/4.0	35.5	38.0	35.5	2.32	3.08	18/20	20/23	13	21
10-27	12.1/4.0	36.2	40.7	36.2	2.05	2.15	15/21	10/21	100	71
10-28	19.7/6.6	40.5	42.5	40.5	2.96	3.37	20/16	5/19	-	44
11-2	20.4/6.8	39.8	39.2	39.2	2.68	3.37	20/16	5/15	46	400
Avg.	17.9/5.9	38.2	37.9	36.6	2.70	3.31	18.4/15.3	10.4/17.1	485	96

FIXED VOLUME FILTER PRESS

The fixed volume press is a standard recessed plate filter press which produces cakes of a constant thickness. These presses are designed to operate at terminal pressures ranging from 7 kg/cm^2 (100 psig) to 15.8 kg/cm^2 (225 psig). For the study, a 225 psig unit was supplied by Passavant Corp., and a 100 psig unit was supplied by Neptune - Nichols Inc. This report refers to the 100 psig press as a "low-pressure unit" and the 225 psig press as a "high-pressure unit". Both presses were operated to develop design data for dewatering a variety of sludge ratios, but most work was centered on the 2/1 secondary to primary ratio. The units were also used for comparison tests with each other and with the diaphragm type press.

High-Pressure Press

Facilities--

The Passavant system included the following equipment.

1. Press - Passavant Model 2400 - Produced up to 6 circular cakes, each 597 mm (23.5 inches) in diameter. Each chamber had a filtration area of 0.56 m^2 (6.0 ft^2). Several stainless steel plates were supplied to provide chamber thicknesses of 30 mm (1.18 inch), 34 mm (1.34 inch), and 38 mm (1.50 inch). A hydraulic mechanism was provided for press closing. (See Figure 19.)

2. Feed tank - 1135 l (300 gal) cylindrical closed tank; capable of withstanding air pressures up to 21 kg/cm^2 (300 psig).

3. Air compressor - operated at pressures up to 21 kg/cm^2 (300 psig) for feeding the press and core blowing.

4. Filtrate collection tank - 378 l (100 gal) calibrated plastic vat.

Operation--

For all runs, the sludge was blended and conditioned in the NGK mix tank prior to pumping to the feed tank. Before each run the cloths were wetted with tap water and scrubbed with a stiff-bristle nylon brush. The press was then closed hydraulically. The sludge feed valve was opened and the high-pressure compressor started. Within 15 to 20 minutes the full pressure of 15.8 kg/cm^2 (225 psig) was attained and was held for the remainder of the run, thus providing the sole driving force for dewatering the sludge. The run was ended when either the filtrate rate reached 0.1 gal/hr/ft^2 or when three hours filtration time had elapsed. Usually 3 cakes were made, but at times when greater quantities of feed sludge were available, up to 5 cakes could be produced. At the end of the run all cakes were weighed and analyzed for percent solids. Generally, the availability of laboratory oven space permitted no more than one or two cake samples for analysis for percent solids. Data sheets 5 and 6 in Appendix B summarize a typical run on the high-pressure press. Explanations are provided with each data sheet.

Test Data--

All test runs with the high pressure-press were conducted in August and September, 1977. During this time, sludge temperatures ranged from 24°C to 30 °C. In August, the feed before conditioning averaged 5.7% solids and the sludge dewaterability was good. In September, the feed averaged from 3.4 to 4.0% solids before conditioning and the dewaterability was poor. Thus, higher chemical dosages were required to filter the sludges than during August.

The data in Table 14 are typical results with this press for a variety of secondary to primary sludge ratios. These tests were all made in August using a 38 mm (1.5 inch) cake. The filter cloth used was a nylon monofilament of twill weave, with an air permeability of 76.7 $cm^3/s/cm^2$ @ΔP= 12.7mm H_2O. The full-scale yields were computed by adding 20 minutes mechanical turn-around time to the process cycle time (based on the manufacturer's recommendation for their largest press). As the ratio of secondary sludge increased, the chemical requirements increased, and the cake solids and yields decreased. Because of the open weave cloth on this press, the filtrate suspended solids were sometimes high, particulary when the sludge was marginally conditioned. The average cake density for the runs was 1123.0 kg/m^3 (70.1 lb/ft^3). Cake discharge from this press was not the best; a thin mat of fibrous sludge remained on the cloth after each run, especially around the center feed hole. No precoat was used for any of these runs.

The cake always showed a very dry outside portion and a much wetter inner section. Because of the limited size of the air compressor receiver, the core blow at the end of each run was generally ineffective in removing all the solids from the center feed hole. During cake sampling, pie shaped pieces were taken which included the proper proportions of this wet inner core and the dry outside section. To insure that these slices were indeed representative, the variation in dryness across the cake was periodically checked. A wedge was divided into four quarters as shown in Figure 20, and each quarter was analyzed for percent solids. These results were:

	% solids	(section)			Theor.	% solids of
Date	I	II	III	IV	% solids	adjacent wedge
8/15	17.4	21.7	37.7	39.3	32.5	31.2
8/16	41.6	47.4	52.1	49.0	48.9	47.5
8/17	19.7	33.0	43.9	42.5	38.5	37.0
8/18	21.6	28.7	41.2	41.9	36.5	36.6
8/19	26.1	35.9	41.7	42.3	39.2	37.4
8/23	16.1	23.4	36.5	40.8	32.8	33.6

Section I was 10.6% of the total volume; section II was 22.8%; section III was 35.1%; and section IV was 31.5%. The theoretical percent solids was calculated by multiplying these percentages by the percent solids in each section. The correlation of the last two columns is quite good, indicating that our method was correct. These results also show that a standard recessed plate press will always have a variation in percent solids across the cake.

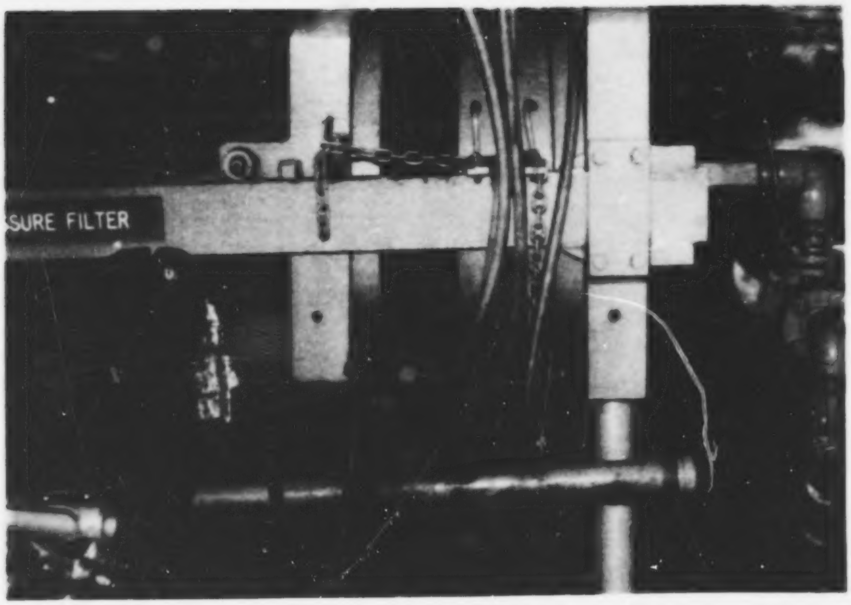


Figure 19. Passavant Filter Press.

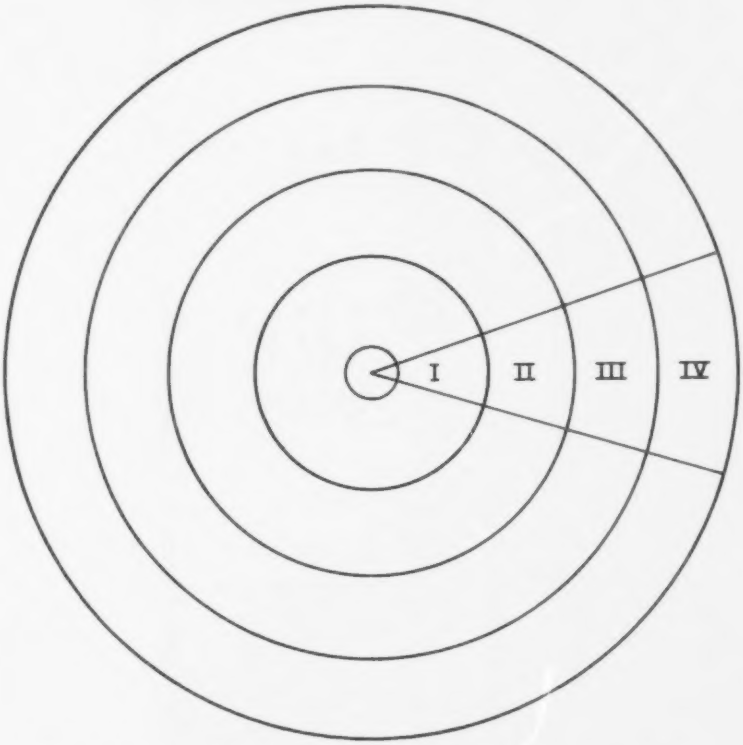


Figure 20. Sample sections from Passavant cake.

TABLE 14. TYPICAL RESULTS ON MODEL 2400 HIGH-PRESSURE (38 mm PLATE) - AUGUST RUNS

SEC/PRIM RATIO	NO. OF RUNS AVERAGED	% CHEMICALS LIME/$FeCl_3$	PROCESS CYCLE TIME (MIN)	% CAKE SOLIDS	FULL-SCALE YIELD ($kg/hr/m^2$)	FILTRATE SOLIDS mg/l TOTAL/SUSPENDED
0/1	2	13.4/4.5	160	47.6	3.37	6305/36
1/2	2	13.2/4.4	125	35.9	2.54	7818/36
1/1	3	15.6/5.2	140	38.6	2.69	7930/37
2/1	13	18.3/6.1	169	34.3	1.94	7610/260
3/1	5	22.7/7.6	146	33.2	2.04	10399/384
4/1	3	21.4/7.2	137	36.9	2.39	7724/43
1/0	4	23.5/7.8	148	29.5	1.70	7770/324

With the more difficult filtering sludges this variation was even greater as shown by the data on 8-15 (100% secondary), and 8-17 through 8-23 (all 2/1 sludge). The 8-16 run on 100% primary showed only slight variation because this sludge was easily filtered.

Table 15 presents the results of runs on the 2/1 secondary to primary sludge. When the sludge was properly conditioned, the high-pressure press gave acceptable cake solids (over 35%), using the 38 mm cake, with cycle times approaching three hours. Marginally conditioned sludges as shown on 8/4, 8/19, and 9/1 gave very poor results for this thick cake. In September, two new plates were installed to provide cakes measuring 30 mm (1.18 inches). The end plates could not be changed, however. So on subsequent runs, the two inside cakes measured 30 mm (1.18 inches), while the two outside cakes measured 34 mm (1.34 inches). When the cakes could be weighed and analyzed separately, a yield was computed for each thickness (runs on 9/21 and 9/22). The runs after 9/14 show that the thinner cakes always contained slightly higher cake solids, but with some measurable sacrifice in overall yield. No comparisons could be made between the runs in August and those in September because the sludge filterability had changed so drastically.

During August, 1977, additional high pressure runs were made on a Passavant Model 600 bench-scale press. This press, 152 mm (6 inches) in diameter, could produce cakes of various thicknesses from 25 mm to 38 mm. The press was fed from a small tank pressurized with nitrogen to 15.8 kg/cm^2 (225 psig). For most runs, the conditioned sludge was sampled from the NGK mix tank. Results are presented in Table 16. In general, the tests showed that higher cake solids were produced by the thinner cakes (see the comparison tests on 8/17, 8/18, and 9/1). Any change in full-scale yield because of the different cake thicknesses was not readily apparent from this data.

Some comparison runs between the Model 600 and Model 2400 presses were conducted for the 38 mm (1.5 inches) cake:

	CYCLE TIME (min)		YIELD (kg/hr/m^2)		% CAKE SOLIDS	
DATE	M-600	M-2400	M-600	M-2400	M-600	M-2400
8-17-77	90	200	-	-	37.8	37.0
8-18-77	120	190	3.12	1.81	35.8	36.6
8-30-77	115	180	2.54	1.47	26.7	28.4
8-30-77	110	180	3.37	1.54	39.2	31.9
9-1-77	130	180	2.83	1.54	35.9	29.3

The above table shows that with the Model 600 press, cycle times were much shorter and the resultant yields were much higher. In some cases higher cake solids were also achieved with this unit. No satisfactory explanation has yet been given for these apparently inconsistent results.

The scale-up factor (filtration area per plate) for the Model 2400 to full-scale is only 12.9 to 1.0, while for the Model 600, it is 198 to 1.

TABLE 15. RUNS ON MODEL 2400 HIGH-PRESSURE PRESS WITH 2/1 SECONDARY/PRIMARY SLUDGE.

DATE	% CHEMICALS LIME/$FeCl_3$	CYCLE TIME (MIN)	% CAKE SOLIDS	CAKE THICKNESS (mm)	CAKE DISCHARGE	FULL SCALE YIELD ($kg/hr/m^2$)
8-2-77	21.9/7.4	120	37.4	38	excellent	2.73
8-2-77	17.9/5.9	150	36.1	38	excellent	2.25
8-4-77	14.7/4.9	120	28.4	38	not noted	2.06
8-17-77	18.0/6.0	200	37.0	38	good	1.81
8-18-77	19.6/6.6	190	36.6	38	good	1.81
8-19-77	21.5/7.2	180	37.4	38	good	1.94
8-19-77	14.2/4.8	180	29.6	38	fair-good	1.60
8-23-77	16.6/5.5	180	33.6	38	fair-good	1.73
8-24-77	18.3/6.2	190	34.1	38	fair	1.63
8-25-77	20.1/6.7	160	36.2	38	not noted	2.26
8-25-77	19.3/6.5	180	35.4	38	good	1.88
9-1-77	17.9/6.0	180	29.3	38	poor	1.54
Average	18.3/6.1	169	34.3	38		1.94
9-13-77	20.1/6.7	180	28.7	30-34	not noted	1.10
9-14-77	24.3/8.1	180	30.1	34		1.27
			33.6	30	good	
9-14-77	29.2/9.7	180	36.2	34		1.48
			37.3	30	good	
9-21-77	27.7/9.2	140	34.4	34		1.85
			36.4	30	good	1.79
9-22-77	30.0/10.0	160	35.9	34		1.67
			36.2	30	good	1.45

TABLE 16. RUNS ON MODEL 600 HIGH-PRESSURE PRESS

DATE	RATIO SEC/PRIM	% CHEMICALS LIME/$FeCl_3$	CYCLE TIME (MIN)	% CAKE SOLIDS	CAKE THICKNESS (mm)	FULL SCALE YIELD ($kg/hr/m^2$)
8-17-77	2/1	18.0/6.0	75	43.1	25	N.A.*
8-17-77	2/1	18.0/6.0	90	37.8	38	N.A.
8-17-77	2/1	17.5/5.8	120	41.7	32	N.A.
8-17-77	2/1	17.5/5.8	120	35.9	38	N.A.
8-18-77	2/1	19.6/6.6	120	35.8	38	3.12
8-18-77	2/1	19.6/6.6	105	40.4	32	3.17
8-23-77	2/1	16.6/5.5	155	40.4	32	2.39
8-24-77	2/1	18.3/6.2	130	39.6	32	2.64
8-24-77	2/1	30.0/10.0	100	32.6	32	2.20
8-25-77	2/1	20.1/6.7	90	24.1	32	2.05
8-30-77	3/1	14.5/4.9	115	26.7	38	2.54
8-30-77	3/1	14.5/4.9	110	27.0	32	2.00
8-30-77	3/1	28.7/9.7	110	39.2	38	3.37
8-30-77	3/1	28.7/9.7	90	40.3	32	3.38
9-1-77	2/1	20.0/6.7	130	39.0	32	2.57
9-1-77	2/1	20.0/6.7	130	33.9	38	2.83

*Not available.

Therefore, Model 2400 data was used for comparisons with the other presses and for full-scale design. As with the other presses, scale-up from pilot data to the full-scale press can be made directly. The process cycle time (filtration time) is assumed to be identical for both size units. The total cycle time, however, must be adjusted to include the mechanical turnaround time in order to obtain the full-scale yield.

Low-Pressure Press

Facilities--

The Nichols system included the following equipment:

1. Press - 0.37 m^2 (4 ft^2) filtration area. Produced two octagon shaped cakes each measuring 330 mm (13 inches) across. Plates were rubber coated steel, with a chamber thickness of 25 mm (1.0 inch). Spacers were available to produce a cake thickness of 32 mm (1.25 inches). The press was closed by a manually operated screw. (See Figure 21.)

2. Feed tank - 113.5 l (30 gal.) cylindrical closed tank; could be pressurized with air up to 10.5 kg/cm^2 (150 psig).

Operation--

For all the runs on this press, the sludge was blended and conditioned in the NGK mix tank prior to pumping to the feed vessel. Before each run, the cloths were wetted with tap water and scrubbed with a stiff-bristle nylon brush. The press was closed and sealed as tightly as possible by manually turning the screw. The inlet valve to the press was opened and the feed tank pressurized slowly with air to reach a pressure of 7 kg/cm^2 (100 psig) within 5 to 10 minutes. This pressure was maintained throughout the entire run and provided the sole driving force for dewatering. Filtrate was collected from a drain pipe and a drip pan under the plates during the run. Because there were no gaskets between the filter plates, and the filter cloths provided the only seals, up to 50% of the filtrate was collected from the drip pan. Early test work established that the run was complete when the filtrate rate reached 25 ml/min or less. At the end of the run both cakes were weighed and sampled. A triangular shaped section, as shown in Figure 22, was taken and analyzed for percent solids. Data Sheets 7 and 8 in Appendix B summarize a typical run on the low-pressure press. Explanations are provided with each data sheet.

Test Data--

Test work with the low-pressure press throughout the year showed that the press could dewater sludges over a range of sludge temperatures from 11 °C to 30 °C. The data in Table 17 are typical results with this press on a variety of secondary to primary sludge ratios with an average unconditioned feed solids concentration of 5.7%. These tests were all conducted in August during comparison studies with the high-pressure unit and the diaphragm press. The sludges tested at that time contained a high proportion of septic solids; however, their dewaterability was quite good even at low chemical dosages. The full-scale yield was computed by adding 20 minutes mechanical turn-around time to the cycle time (based on the manufacturer's recommendation for the largest press available). Attempts to run 100% primary sludge on this small pilot press failed because solids

Figure 21. Nichols Filter Press.

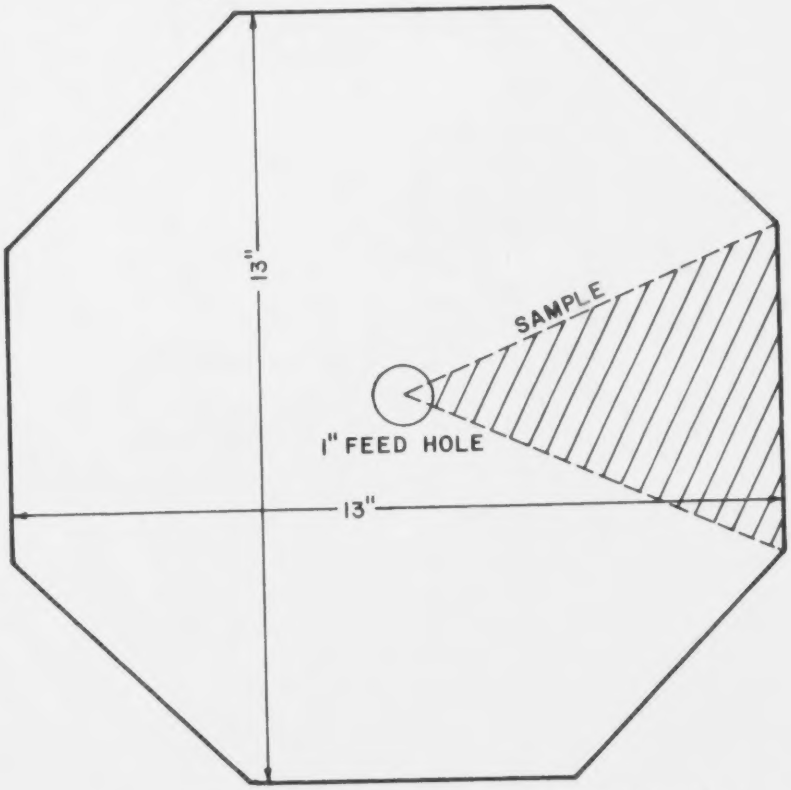


Figure 22. Sample sections from Nichols cake.

TABLE 17. TYPICAL RESULTS ON LOW PRESSURE PRESS - AUGUST RUNS

SEC/PRIM RATIO	NO. OF RUNS	% CHEMICALS LIME/$FeCl_3$	PROCESS CYCLE TIME (MIN)	% CAKE SOLIDS	FULL-SCALE YIELD(kg/hr/m^2)	FILTRATE SOLIDS mg/1 TOTAL/SUSPENDED
1/2	2	13.2/4.4	128	38.5	1.61	7382/44
1/1	3	15.6/5.2	127	39.0	1.59	8532/57
2/1	13	18.3/6.1	140	35.0	1.39	8082/50
3/1	5	22.7/7.6	140	34.1	1.32	10267/69
4/1	3	21.4/7.1	113	35.6	1.46	7786/36
1/0	3	23.1/7.7	143	32.3	1.08	7125/49

plugged the small (1 inch diameter) feed line.

The general trend, as shown in Table 17, was that as the ratio of secondary sludge increased, the chemical requirements increased, and the cake solids and yields decreased. The exception to this was the 4/1 sludges which seemed to filter extremely well. The filtrate quality was unaffected by changes in the sludge ratios. The cloth used for all these runs was the Nichols 4709/40 cloth (a monofilament fabric with a 2 x 2 twill weave, and an air permeability of 20.3 $cm^3/s/cm^2$ @$\Delta P = 12.7$ mm H_2O). The average cake density was 1141 kg/m^3 (71.2 lb/ft^3). Except for the marginally conditioned sludges, cake release from the cloth was generally very good. But, as with the high-pressure units, the cake from this press was drier on the outer sections than at the inner core.

Table 18 presents the results of the 13 individual runs on the 2/1 secondary/primary sludge. The first three runs in the table, 8/2/77 and 8/4/77, show the effect of chemical conditioning. Low chemical dosages (marginal conditioning) as in the run on 8/4/77 gave poor results on this press. The first eight runs in the table were the results with the 25 mm (1 inch) thick cake; the last 5 were with the 32 mm (1.25 inches) thick cake. Increasing the cake thickness to 32 mm (1.25 inches) provided some tradeoffs. The averages of the two sets of runs showed that resultant cake solids were slightly lower with the thicker cake, but the overall full-scale yields were nearly identical at approximately 1.4 kg/hr m^2 (0.29 lb/hr ft^2). The increased cake thickness also required increased cycle times.

Comparison Runs

The comparison runs in August, 1977, were designed to establish the operating conditions for the three types of presses-- low-pressure fixed volume, high-pressure fixed volume, and diaphragm. These tests were run on seven different secondary to primary sludge ratios. With each ratio, at least three runs were made: one with the sludge over-conditioned; one with the sludge conditioned in a good, safe range; and one with the sludge marginally conditioned. The lime/$FeCl_3$ dosages required to produce the above conditions were determined from experience; filterability was checked by specific resistance and CST tests prior to each run.

Facilities--

1. Diaphragm press - NGK unit 5.8 m^2 (62.4 ft^2) filtration area.
2. High-pressure fixed volume - Passavant Model 2400 with 1.67 m^2 (18 ft^2) filtration area (3 cakes). All tests used the 38 mm plates.
3. Low-Pressure fixed volume - Nichols unit - .37 m^2 (4 ft^2) filtration area. Cake thickness was either 25 mm or 32 mm.

Operation--

Thickened primary and secondary sludges were independently pumped to the NGK mix tank and blended in the proper proportions for the test. Percent solids of each sludge were determined prior to blending to be certain that

TABLE 18. RUNS ON LOW-PRESSURE PRESS WITH 2/1 SECONDARY/PRIMARY SLUDGE

DATE	% CHEMICALS LIME/$FeCl_3$	PROCESS CYCLE TIME (MIN)	% CAKE SOLIDS	CAKE THICKNESS (mm)	CAKE DISCHARGE	FULL SCALE YIELD (kg/hr/m^2)
8-2-77	22.1/7.4	100	37.5	25	ecxellent	1.71
8-2-77	17.9/5.9	120	37.5	25	excellent	1.51
8-4-77	14.7/4.9	140	31.0	25	fair - good	1.12
8-17-77	18.0/6.0	110	35.3	25	excellent	1.51
8-19-77	21.5/7.2	140	35.1	25	excellent	1.17
8-19-77	14.2/4.8	110	34.1	25	excellent	1.56
8-25-77	20.1/6.7	130	36.2	25	excellent	1.32
9-1-77	17.9/6.0	150	35.8	25	good	1.22
Average	18.3/6.1	125	35.3	25		1.39
8-17-77	17.5/5.8	180	34.4	32	good	1.27
8-18-77	19.6/6.6	150	34.5	32	excellent	1.51
8-23-77	16.6/5.5	180	32.5	32	good	1.22
8-24-77	18.3/6.2	170	34.0	32	excellent	1.32
8-25-77	19.3/6.5	140	36.4	32	good	1.66
Average	18.3/6.1	164	34.4	32		1.40

the blend was accurate. $FeCl_3$ solution at one lb/gal was added and allowed to mix in. Lime solution at one lb/gal was then added and mixed in. Within 15 minutes after adding the lime, portions of the conditioned sludge were pumped to the Passavant and Nichols feed tanks. The remainder was fed to the NGK press. The three presses were all started at the same time. The operational procedures, (i.e. cycle times, pressures, etc.) were those described previously.

Test Data--

Individual test runs on the three presses are presented in Table 19. All runs on the 2/1 sludges are included, while only the best runs with average conditioning are presented for the other sludge ratios. The full-scale yields were calculated with these turn-around times: NGK - 19 minutes, Passavant - 20 minutes, and Nichols - 20 minutes. Several important conclusions were derived from this table:

1. With a properly conditioned sludge, all three types of presses proproduced the required 35% cake solids.

2. On the average, the diaphragm press gave both higher cake solids (40.0% vs 34.3% and 34.9%) and higher yields than the fixed volume presses. Cake solids were approximately the same on both the high and low pressure presses, but the high-pressure press gave significantly greater yields than the low-pressure unit.

3. The diaphragm press was the only unit capable of satisfactorily dewatering the marginally conditioned sludges. For example, on 8/4 and 8/30 both the Passavant and Nichols presses had poor runs, with wet, sloppy cakes and extremely low yields. With the same sludge the diaphragm press gave a good cake discharge and high cake solids, but at reduced yields. The diaphragm press, because of its separate squeezing cycle, provided a much more flexible operation. These poorly conditioned sludges were pumped for shorter cycles and the squeezing time was increased slightly to give thin, dry cakes. The fixed-volume presses did not have this option, so once the sludge was fed to these presses no corrective measures could be taken.

4. The marginally conditioned runs on 8/19 and 9/1 gave poor cake solids on the Passavant press, but acceptable results on the Nichols press. This indicated that cake thickness had more of an effect than pressure in determining cake solids content.

5. As the percentage of primary sludge increased, the cake solids and yields improved for all presses. Cake solids approached the 50% solids level for high primary ratios.

6. For the properly conditioned 2/1 sludge runs the average filtrate suspended solids were: NGK - 197 mg/l; Passavant - 26.5 mg/l; Nichols - 49.1 mg/l. The calculated percent recovery of inlet suspended solids in the filter cake is: NGK - 99.74%; Passavant - 99.97%; and Nichols - 99.93%.

TABLE 19. COMPARISON RUNS

DATE	RATIO SEC/PRIM	% CHEMICALS LIME/FeCl$_3$	% CAKE SOLIDS NGK	% CAKE SOLIDS PASS	% CAKE SOLIDS NICHOLS	FULL-SCALE YIELD (kg/hr/m^2) NGK	FULL-SCALE YIELD (kg/hr/m^2) PASS	FULL-SCALE YIELD (kg/hr/m^2) NICHOLS
8-2-77	2/1	22.1/7.4	41.4	37.4	37.5	3.06	2.73	1.71
8-2-77	2/1	17.9/5.9	40.8	36.1	37.5	2.83	2.25	1.51
8-4-77	2/1	14.7/4.9	40.2	28.4	31.0	2.17	2.06	1.12
8-17-77	2/1	18.0/6.0	39.9	37.0	35.3	2.90	1.81	1.51
8-17-77	2/1	17.5/5.8	39.1	-	34.4	2.69	-	1.27
8-18-77	2/1	19.6/6.6	40.2	36.6	34.5	2.77	1.81	1.51
8-19-77	2/1	21.5/7.2	42.2	37.4	35.1	3.19	1.94	1.17
8-19-77	2/1	14.2/4.8	36.0	29.6	34.1	2.12	1.60	1.56
8-23-77	2/1	16.6/5.5	36.4	33.6	32.5	2.13	1.73	1.22
8-24-77	2/1	18.3/6.2	41.4	34.1	34.0	2.19	1.63	1.32
8-25-77	2/1	20.1/6.7	43.0	36.2	36.2	3.07	2.26	1.32
8-25-77	2/1	19.3/6.5	40.9	35.4	36.4	2.77	1.88	1.66
9-1-77	2/1	17.9/6.0	38.8	29.3	35.8	2.34	1.54	1.22
Average	2/1	18.3/6.1	40.0	34.3	34.9	2.63	1.94	1.39
8-5-77	3/1	25.5/8.5	42.4	37.0	35.4	2.52	2.33	1.27
8-30-77	3/1	14.5/4.9	37.2	28.4	29.3	1.80	1.47	1.07
8-8-77	4/1	19.5/6.5	43.4	38.8	35.6	3.25	2.38	1.46
8-10-77	1/1	15.2/5.0	47.0	38.2	39.6	3.64	2.53	1.95
8-11-77	1/2	15.5/5.2	48.2	40.7	39.0	4.17	2.74	1.90
8-16-77	0/1	15.7/5.3	55.2	47.5	-	4.92	4.42	-
8-15-77	1/0	28.1/9.3	41.4	31.2	32.6	2.72	1.66	1.32

Overall, the diaphragm press performed better than either of the fixed volume presses. But since these presses also produced the required cake solids with the 2/1 sludge, the comparison for cost and design purposes was based on the respective yields achieved in reaching the 35% solids cake. The runs that reached the required cake solids for the low and high-pressure presses are shown in Figure 23. Full-scale yields are plotted vs the date of the run and the eight points are averaged for each unit. The NGK cycle time was then adjusted to a level which would give the same cake solids as obtained on the other presses. This was done by recalculating the squeezing time to the point where the desired percent solids were reached. For example, the first NGK run on 8/2 had a pump/squeeze time of 21/16 minutes and a final solids of 41.4%. The Passavant and Nichols cake solids were 37.4%. Filtrate collection data for the NGK run showed that this 37.4% solids level was reached after 9 minutes of squeezing. Therefore, a recalculated NGK cycle time of 21 minutes pumping, 9 minutes squeezing, and 19 minutes mechanical time was used, and a full-scale yield of 2.94 $kg/hr/m^2$ obtained. These adjusted NGK yields are also plotted in Figure 23 and the eight runs averaged. To produce a 36.3% solids cake the NGK average yield was 3.31 $kg/hr/m^2$; the Passavant yield was 2.04 $kg/hr/m^2$; and the Nichols yield (for 35.8% solids) was 1.46 $kg/hr/m^2$.

Using the above yield data, the filtration area required to process a given quantity of sludge was then computed for each of the press types. For example, to process 1000 kg/hr of dry sludge solids, the NGK press would have required:

$$\frac{1000 \text{ kg/hr}}{3.31 \text{ kg/hr/m}^2} = 302 \text{ m}^2 \text{ of filtration area.}$$

Likewise, the Passavant and Nichols presses would have required 490 m^2 and 685 m^2 of filtration area, respectively. Thus using the NGK press as a base, the Passavant unit requires 62.3% more filter area, and the Nichols press requires 126.8% more filter area than the NGK press to dewater the same quantity of sludge. It must be noted that this relationship was derived specifically for the Blue Plains sludge. It is further noted that the relative filter areas refer only to the largest press sizes available from each of the manufacturers: NGK - 500 m^2; Passavant - 1080 m^2; and Nichols - 628 m^2. For comparisons of smaller size presses, a different mechanical time must be used and a new full-scale yield must be calculated for each unit. For convenience, the following table shows the process cycle time used to derive Figure 23:

Process Cycle Times
(min)

Date	NGK	Pass.	Nich.
8-2	21/9	120	100
8-2	20/9	150	120
8-17	16/11	200	110
8-18	18/11	190	150
8-19	20/8	180	140
8-24	21/12	190	170
8-25	18/7	160	130
8-25	17/10	180	140

See Table F-1 in Appendix F for specifications of the large-scale presses available from each of the manufacturers.

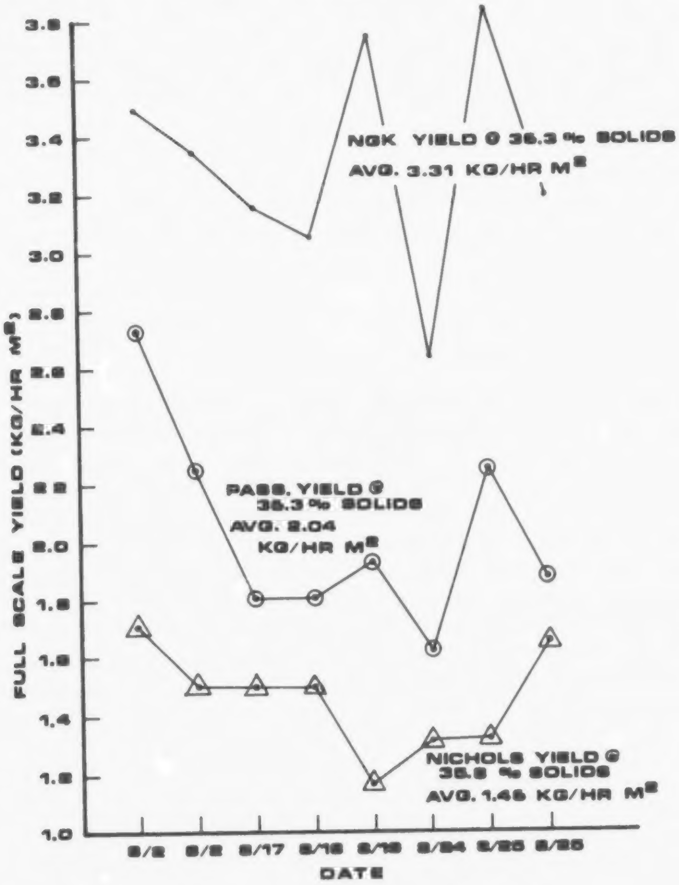


Figure 23. Comparative yield data.

CONTINUOUS BELT FILTER PRESSES

Continuous belt presses for sludge dewatering were originally developed in Europe and have generally found wide acceptance in many countries. Several companies in the United States have purchased the technology and are now marketing these presses in this country. Two manufacturers' units were tested in the study - a Parkson Magnum press and a Komline Sanderson Unimat press. The units were used to dewater both thickened sludge and cake from a vacuum filter.

Parkson Magnum Press

The Magnum press was equipped with two continuous screens made of polyester monofilament cloth (air permeability of 228 $cm^3/sec/cm^2$ @$\Delta P = 12.5$ mm water) which ran through a system of guiding and pressing rollers that were perforated to allow for water drainage. There were three dewatering zones; a gravity drainage stage, a low-pressure stage - 0.5 kg/cm^2 (7.5 psig), and a high-pressure stage - up to 7.7 kg/cm^2 (110 psig). Their full-scale press has a range of belt speeds from 1.2 to 5.7 m/min. (See Figure 24.)

Facilities--

A 0.25 meter wide laboratory press was tested in May, 1977. A 1.0 meter wide trailer-mounted demonstration unit was tested in October, 1977. A hopper and Moyno pump were supplied with the demonstration unit to test the vacuum filter cake as feed.

Operation--

The laboratory unit (0.25 meter), pictured in Figure 25, was used to provide basic information on dewatering polymer conditioned thickened sludge blends. Data was also collected for the further dewatering of vacuum filter cake. Various ratios of secondary to primary sludge were blended in laboratory glassware, and the polymer was added and mixed in. The conditioned sludge was placed on the drainage section of the belt press and, after a suitable drainage time, the belts moved the sludge through the pressure zones. Yield was computed from measurements of cake weight and belt speed. Solids recovery on this unit was estimated from experience. When testing vacuum filter cake, samples were taken from the plant's full-scale units and manually placed on the press.

When testing on the demonstration size press, vacuum filter cake was collected from the full-scale filters in a truck, dumped on the ground, and then loaded in the feed hopper with a front end loader. An open-throat Moyno pump fed the sludge through a six-inch hose and a variable orifice feed nozzle onto the drainage section of the press.

Test Data--

Laboratory (0.25 meter) unit--Results of the tests with varying ratios of secondary to primary sludge are presented in Figures 26 and 27. With a feed range of 5.5% to 9.5% total solids, the final cake solids increased linearly from 25% to 41% as the percentage of primary increased. The press

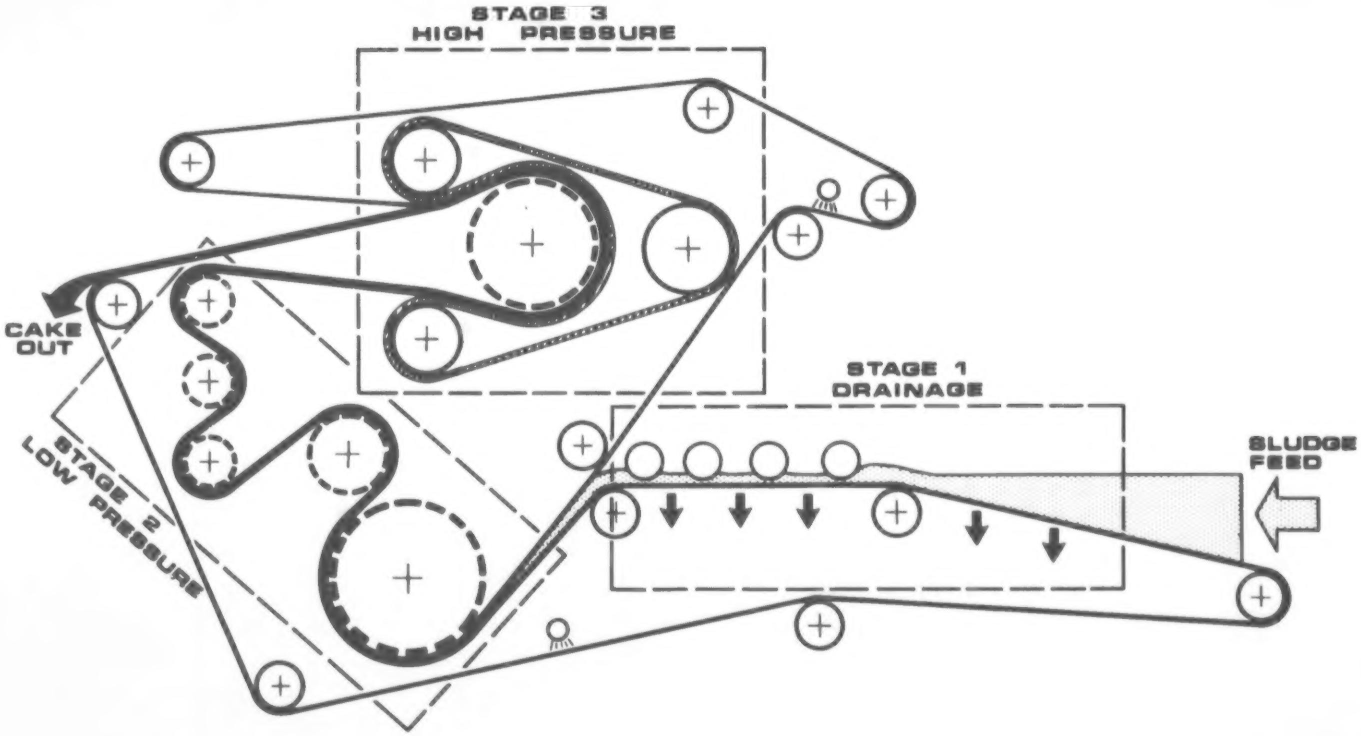


Figure 24. Schematic of Parkson belt press.

Figure 25. Parkson Laboratory Belt Press.

capacity exhibited an S-shaped curve, ranging from 248 to 1230 kg/hr/meter of belt width (547 to 2712 lbs/hr) for pure secondary and pure primary, respectively. Figure 27 shows that the polymer (Percol 721 @ $1.70/lb) consumption decreased from 5.5 to 1.6 lbs per ton dry solids, and estimated solids recovery (reflecting losses in both filtrate and washwater) increased from 95% to 98% with increasing percent primary. With high primary sludge (greater than 84% by weight) the belt speeds were near maximum of 5 meters/min and high pressures of 7 kg/cm^2 were attained. As the percent primary decreased, the belt speed was reduced to 3 meters/min and pressures of only 1.8 kg/cm^2 (25 psig) were applied. With the high primary sludges, cake release from the cloth was excellent as is pictured in Figure 25. As the percent primary decreased, a sharp scraper blade was needed to remove the cake. Some solids, however, usually remained imbedded in the cloth and high pressure washing was required to remove them.

The tests with vacuum filter cake showed excellent results. At that time, average conditioning chemicals of 19% lime, 6% $FeCl_3$, and 0.14% polymer were added to the vacuum filter feed. Cake solids from the vacuum filter averaged 20%; no additional chemicals were mixed with the belt press feed. The following table shows how the cake solids varied with throughput rate:

Capacity (kg total solids/hr/m)	378	702	972	1260
% Cake Solids	42	39	36	35

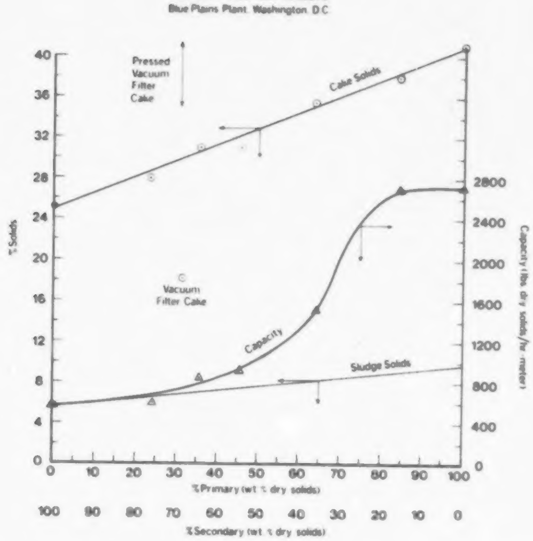


Figure 26. Results of tests with varying ratios of secondary to primary sludge (Parkson Corporation).

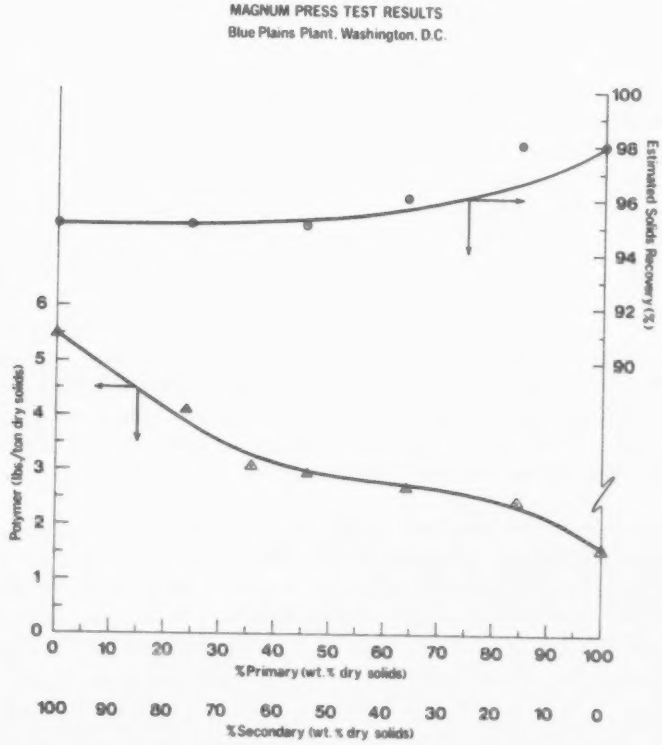


Figure 27. Polymer dosage and solids recovery for varying ratios of secondary to primary sludge (Parkson Corporation).

Solids retention on the belt was estimated at 99%. Cake release was excellent, similar to the release with 100% primary sludge. These tests clearly indicated that a belt press retrofitted to a vacuum filter could produce cake solids in the desired auto-combustible range. Further tests were therefore conducted on a full-scale unit.

Demonstration (1.0 meter) unit--Vacuum filter cake tests on the full-scale unit encountered difficulty and the good results with vacuum filter cake on the laboratory unit were not duplicated. Test results are presented in Table 20. All the problems were related to the feeding and distribution of the vacuum filter cake (cake solids at 20%). In Runs #1 through #3 in Table 20, cake directly from the vacuum filter was used. The feed system was that previously described. The sticky nature of the sludge caused it to hang up on the walls of the hopper and form a bridge across the pump inlet. Some wash water was added to the hopper to facilitate feeding the pump, but interruptions of flow to the press were numerous, and the cake had to be forced manually into the bottom of the hopper. In Run #1, the feed layer was too thick and a shearing and rolling effect at the beginning of the high-pressure section resulted. This condition caused the screen to wrinkle and crease under pressure. In Run #3, the pressure was lowered to a point (approximately 2.1 kg/cm^2) where the material would not extrude from the sides at the high pressure roller. Because of these feed problems, the yields were low in these first three runs. Filtrate suspended solids were measured at 1328 mg/l in Run #1, thus giving only a 95% solids retention on the press.

TABLE 20. PARKSON PRESS AS A RETROFIT TO VACUUM FILTERS

RUN NO.	BELT SPEED (m/min)	HIGH PRESSURE (kg/cm^2)	% CAKE SOLIDS	CAPACITY TOTAL SOLIDS (kg/hr/m width)	REMARKS
1	2	4.4	35.5	287	Matl. directly from filter
2	3	3.9	35.8	323	
3	3	2.1	35.1	376	
4	3	2.3	28.0	339	Matl. from screw conveyor
5	3	1.8	29.4	316	
6	3	2.1	29.0	316	
7	3	2.1	30.6	303	
8	1	2.3	35.1	115	

The vacuum filter cake was then processed through a screw feeder in order to make the cake more fluid. This material was easily fed through the hopper/pump arrangement. Runs #4 through #8 show that the material, although more fluid, also became more difficult to press as the floc deteriorated with the screw action. The speed on the press had to be reduced by one-third in order to achieve the 35% cake solids. Filtrate suspended solids increased to approximately 1900 mg/l and solids retention in the press was only 93%. Much more work in developing an acceptable feed system is required in order to use the belt press in this application.

Komline-Sanderson Unimat Belt Press

Facilities--

The Unimat press was similar in concept to the Parkson press. The Unimat press, pictured in Figure 28, had four dewatering zones: a gravity drainage stage; and low, medium, and high pressure stages. Pressures in the high-pressure section were in excess of 2.1 kg/cm^2 (30 psig). The trailer mounted unit tested was their GM$_2$H - 5/7 pilot plant model, with an effective width of 0.5 meter.

Operation--

For the thickened sludge tests, the primary and secondary sludges were thickened separately and blended in an 11.4 m^3 (3000 gal) tank to produce a 2/1 secondary to primary sludge solids ratio. The blended sludge was metered to a flocculation tank and polymer was added prior to feeding the press. The final cake and filtrate were analyzed for total solids. Yield was determined from measurements of the total solids and flow rate of the feed.

For the vacuum filter cake tests, a truck-load of 20% solids cake was taken from the plant's full-scale units and delivered to a point adjacent to the trailer. The cake was manually fed to the low-pressure section of the belt press in bucket loads. The yield was estimated by counting buckets per unit time. During these tests with the vacuum filter cake, problems were encountered with the motor drive on the press. At times, the motor was overloaded and kept kicking out; a slightly larger motor and drive probably should have been used to handle this feed.

Test Data--

Thickened sludge feed--(2/1 secondary/primary). Table 21 summarizes the results of the tests run with the thickened sludge. During this test period, the plant experienced some upset conditions and the sludge was septic when received. The sludge characteristics varied considerably from days when the polymer would not flocculate the sludge to days when the same polymer worked very well. Thus, the results were quite inconsistent. Laboratory tests were made each morning to determine which of the two available polymers would work. These results show the range of cake solids that were achieved, depending on whether the sludge would respond to the polymer at that time. The first two runs on 7/26 show that a doubling of the polymer rate had only marginal results on the final cake solids. Unlike other chemical conditioning agents, the polymers appeared to be quite selective and worked only within a very narrow range. The last three runs on 7/26 were the best for the entire series, and seem to be representative of what the belt press can produce with the proper polymer conditioning. The overall average results were cake solids of 31 to 33% at a rate of 307 kg/hr/meter of belt width with a polymer cost of approximately $9.00 per ton of sludge solids. Unfortunately, because the Blue Plains sludge is so variable, these results would not be obtained every day. With our type of sludge, a number of polymers would have to be readily available for

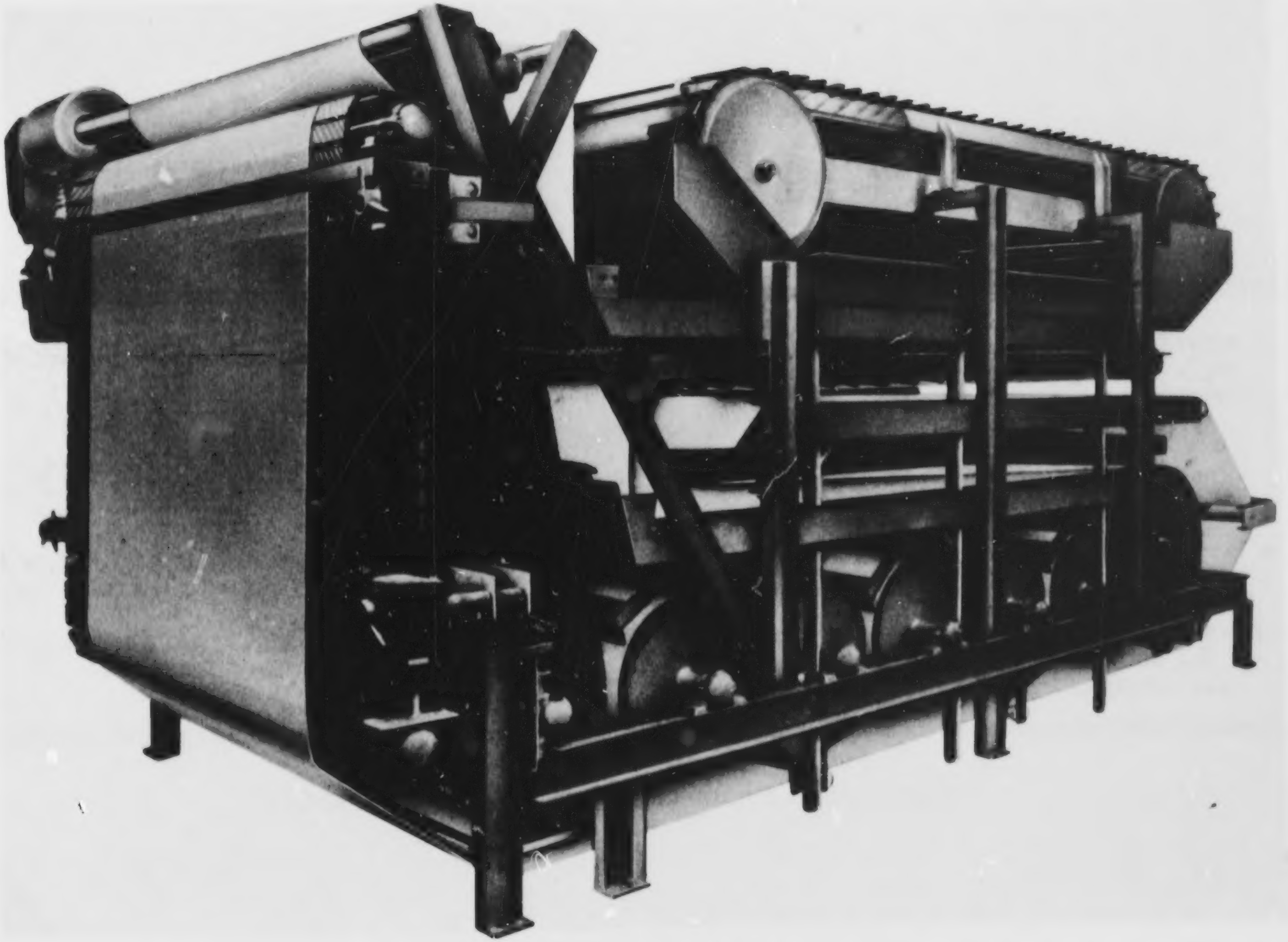


Figure 28. Unimat Belt Press.

TABLE 21. UNIMAT BELT PRESS RESULTS ON 2/1 SLUDGE

DATE	% FEED SOLIDS	POLYMER lb/ton D.S.	POLYMER $/ton	% CAKE SOLIDS	YIELD kg/hr/meter width	FILTRATE *** SOLIDS (mg/l) TOTAL/SUSPENDED
7-19-77	7.1	7.46*	$11.19	24.3	370	1700/ -
7-20-77	6.2	73.0**	$ 9.49	27.6	385	1800/ -
7-21-77	5.2	68.7**	$ 8.93	22.7	322	300/ -
7-26-77	4.5	70.8**	$ 9.20	27.3	292	900/ -
7-26-77	4.5	136 **	$17.68	29.2	292	- / -
7-26-77	5.2	5.84*	$ 8.76	31.3	307	880/384
7-26-77	5.2	5.84*	$ 8.76	32.8	307	880/384
7-26-77	5.2	5.84*	$ 8.76	31.9	307	880/384

* Percol 776

** Calgon 2820

*** includes filtrate and wash water

immediate use as the sludge characteristics changed. Washwater flow rates were not measured during any of these tests, therefore solids recovery was not computed.

TABLE 22. UNIMAT PRESS AS A RETROFIT TO VACUUM FILTER

DATE	% FEED SOLIDS	% CAKE SOLIDS	TOTAL SOLIDS YIELD kg/hr/meter width	FILTRATE SOLIDS mg/l TOTAL/SUSPENDED
7-13-77	22.1	34.3	655	- / -
7-13-77	22.3	35.5	595	- / -
7-13-77	22.3	30.4	947	2800 / -
7-14-77	22.0	37.6	543	- / -
7-14-77	22.8	35.3	585	- / -
7-25-77	23.0	33.2	613	1776 / 1244
7-25-77	23.6	34.1	1399	- / -

Vacuum filter cake--Table 22 shows results for the vacuum filter cake feed. Because the vacuum filter cake was originally conditioned with lime and ferric chloride, and because it was hand fed to the press, these results are consistent. The runs on 7/13 show that as the yield (i.e. belt speed) was increased, the final cake solids decreased. Cake samples taken from the intermediate-pressure zones during these runs showed that the first two zones increased the cake solids from 22% to approximately 27%. The high-pressure zone, on the other hand, increased cake solids from about 27% to approximately 35% and was responsible for the majority of the dewatering. These results look promising and indicate that further tests, perhaps using only a high-pressure section of the press, are warranted.

VACUUM FILTER RETROFIT - ENVIROTECH HI-SOLIDS FILTER

Envirotech Corporation has developed a retrofit unit for a belt-type vacuum filter. This "Hi-Solids" filter is an expression device which extracts additional moisture from a vacuum filter cake. It is equipped with an air compressor and associated controls, in addition to the standard auxiliary equipment needed for a vacuum filter. The unit is pictured in Figure 29. The filter cloth leaves the drum at its uppermost point and travels over a stationary grid. Above the grid is a rubber diaphragm which applies pressure to the cloth and the filter cake. A vacuum is pulled on the bottom of the grid to carry away extracted moisture. The operation of the unit is on a discontinuous cycle. A typical cycle takes either 5.2 or 7.2 minutes per revolution, corresponding to a 20 second or 40 second press time. For a 20 second press time the following sequence occurs: the cake forms on the drum for 20 seconds, the drum progresses 1/5 a revolution

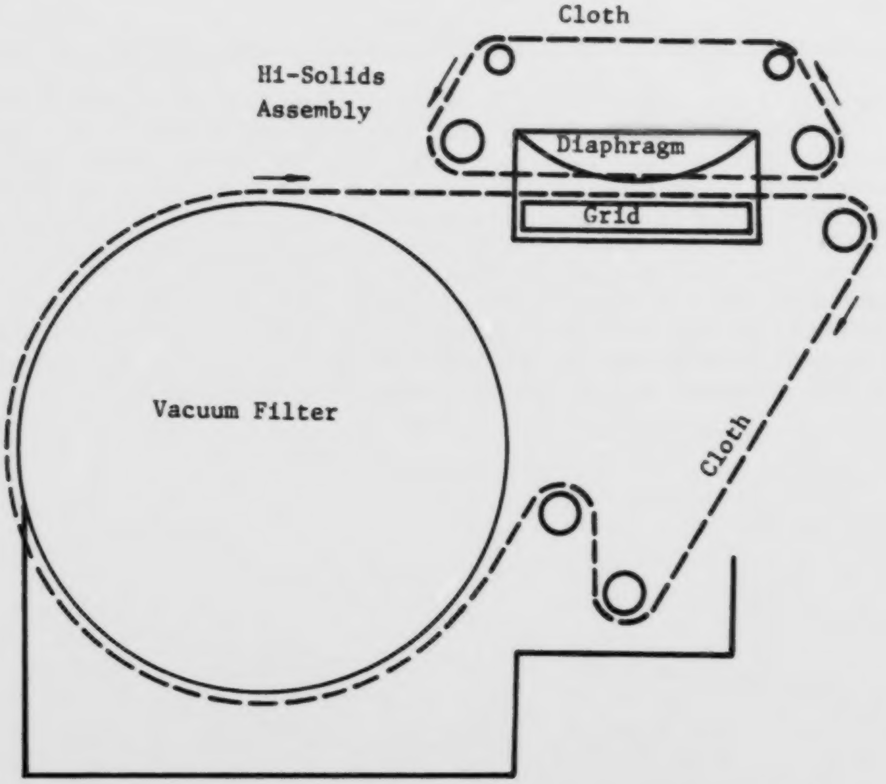


Figure 29. Schematic of Envirotech Hi-Solids Filter

and stops; the cake dries under vacuum for 20 seconds and the drum progresses another 1/5 revolution, etc. When the cake reaches the press zone, it is squeezed by the diaphragm for 20 seconds; then it progresses 1/5 revolution and is discharged. The cloth is subsequently washed and the cycle repeated. Diaphragm pressure is a maximum of 10.5 kg/cm^2 (150 psig).

Facilities--

Test work was conducted in April, 1976 on Envirotech's trailer-mounted, 3 ft. diameter by 3 ft. face filter. Auxiliary equipment included a sludge feed tank, three chemical feed tanks, flocculator, air compressor, transfer pumps, vacuum pump, filtrate pump and receiver.

Operation--

A vacuum filter leaf apparatus was used to evaluate the proper filter media and optimum chemical dosage for each of six different sludge ratios. Candidate filter media were evaluated by conducting five consecutive top feed leaf tests. Selection was based on good cake discharge, filtrate quality, and apparent resistance to cloth blinding. Chemical dosages were optimized by determining the minimum concentrations which gave the maximum filtrate volume for the total dewatering time.

For the tests on the Hi-Solids Filter, primary sludge was obtained from a pilot plant primary clarifier, and secondary sludge was obtained from the plant's secondary clarifiers. Both sludges were thickened and delivered to the trailer feed tank for blending. Using the predetermined chemical dosages, the Hi-Solids Filter was operated at two different test conditions for each sludge:

Condition	Press Time	Cycle Time
1	20 seconds	5.2 min/rev.
2	40 seconds	7.2 min/rev.

Test Data--

The results of the chemical conditioning tests are plotted vs percent secondary sludge in Figure 30. Using these chemical dosages for each sludge ratio, performance tests on the Hi-Solids Filter were conducted. The results are presented in Table 23. Once again the results show decreasing cake solids and yields with increasing ratio of secondary sludge.

The 2/1 ratio was used for comparison with a standard vacuum filter installation. Review of the data showed that a vacuum filter gave a 17% solids cake at 5.2 MPR, and 17.5% solids at 7.2 MPR cycle time. The Hi-Solids Filter increased this to 24 and 25% solids, but was not able to achieve the desired 35% cake solids. Full-scale yield on a vacuum filter with optimum chemical conditioning is expected to be 14.6 kg/hr/m^2. Because of the Hi-Solids Filter attachment, this yield was not achieved at the 20 second press time (5.2 MPR) (14.0 kg/hr/m^2) and was even further reduced (11.3 kg/hr/m^2) with the 40 second press time (7.2 MPR). Because cake solids above the 25% range were not produced, the Hi-Solids Filter was not considered as a dewatering option for the plant.

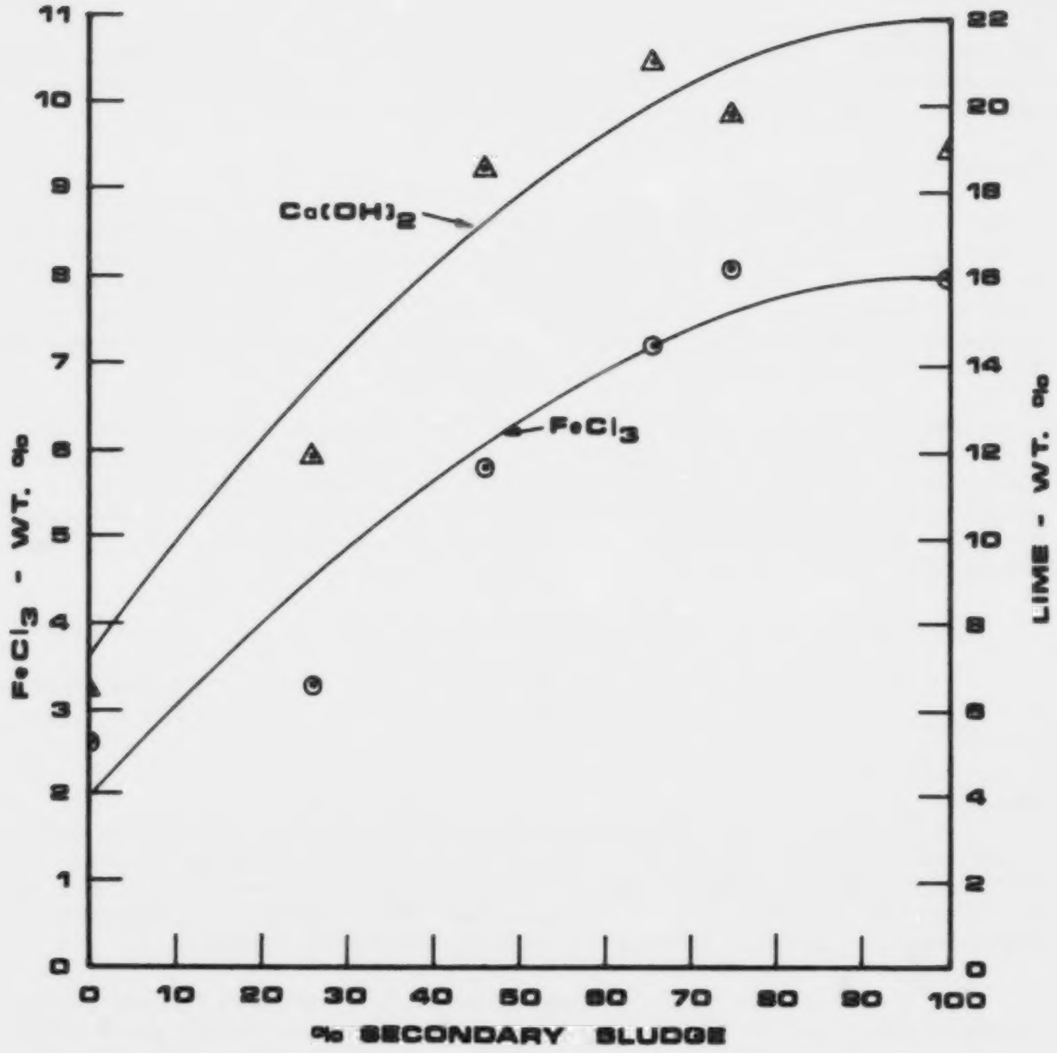


Figure 30. Chemical dosages vs percent secondary sludge. Envirotech tests.

TABLE 23. HI-SOLIDS FILTER RESULTS

RATIO SEC/PRIM	% FEED SOLIDS**	% CHEMICALS LIME/$FeCl_3$	% CAKE SOLIDS 5.2 MPR**	% CAKE SOLIDS 7.2 MPR	FULL-SCALE YIELD kg/hr/m^2 5.2 MPR	FULL-SCALE YIELD kg/hr/m^2 7.2 MPR
0/1	6.9	6.5/2.6	40.7	N.R.	25.2	20.7
1/3	5.4	11.9/3.3	31.0	31.6	21.2	17.7
1/1	4.9	18.5/5.8	25.9	27.1	13.0	10.4
2/1	3.9	21.0/7.2	24.0	25.0	14.0	11.3
3/1	4.1	19.8/8.1	23.8	27.0	14.3	11.6
1/0	3.9	19.0/8.0	N.R.***	N.R.	10.1	8.4

* before conditioning
** MPR = minutes per revolution
*** N.R. = not reported
Vacuum Level for all tests-20 in. Hg
Submergence for all tests - 15%

VACUUM FILTER

During the August, 1977 comparison runs on the filter presses, samples of the conditioned sludge were also used to run a series of optimizing tests on a vacuum filter leaf. Tests were conducted according to the procedure described in the Komline-Sanderson Engineering Corporation instruction manual entitled "Test Leaf Instructions - Rotary Drum Vacuum Filter" (Document Number KSM029). The 0.1 ft^2 leaf, when used properly, can give excellent correlations with actual full-scale vacuum filter operation.

For each sludge sample, at least five different form and dry times were run to show a range of possible operating conditions. The best of each of these runs is shown and compared to the corresponding NGK run in Table 24. Generally, vacuum filter performance was adversely affected by high ratios of secondary sludge; cake solids above 20% were easily achieved when the percentage of secondary sludge was 50% or less. The filter press usually gave approximately twice the percent solids achieved with a vacuum filter; however, because of its continuous operation, the vacuum filter yields were much higher than the filter press yields (i.e. the total filtration area required was much less for a vacuum filter). In all cases, and especially with the higher percentages of secondary sludge, the filter press could more easily dewater varying sludge feeds. Marginally conditioned sludges, or the difficult 100% secondary sludges, gave poor results on the vacuum filter but gave acceptable yields and cake solids on the filter press.

TABLE 24. COMPARISON RUNS - VACUUM FILTER/FILTER PRESS

DATE	RATIO SEC/PRIM	% CHEMICALS LIME/$FeCl_3$	% CAKE SOLIDS		FULL-SCALE YIELD kg/hr/m^2	
			Vac. Filter	NGK Press	Vac. Filter	NGK Press
8-11-77	1/2	10.9/3.6	23.5	47.9	19.5	3.06
8-9-77	1/1	19.6/6.5	19.4	46.8	25.9	3.40
8-10-77	1/1	15.1/5.0	21.4	47.0	22.5	3.64
8-10-77	1/1	12.1/4.0	21.5	48.5	21.0	3.35
7-27-77	2/1	26.8/8.9	22.2	44.9	13.2	3.71
8-2-77	2/1	22.1/7.4	17.1	41.4	7.8	3.06
8-4-77	2/1	14.7/4.5	21.2	40.2	13.2	2.84
8-4-77	3/1	25.8/8.6	18.6	40.6	40.0	2.53
8-5-77	3/1	18.8/6.3	21.8	38.0	18.6	2.24
8-5-77	3/1	25.5/8.3	19.6	42.4	27.3	2.52
8-8-77	4/1	29.6/9.9	18.2	44.9	25.4	3.36
8-8-77	4/1	19.5/6.5	19.5	43.4	38.6	3.25
8-9-77	4/1	15.2/5.1	18.5	41.0	6.8	2.57
8-3-77	1/0	22.9/7.6	14.2	39.2	7.3	1.87
8-3-77	1/0	18.4/6.1	12.7	37.6	4.4	2.79

SECTION 7

SPECIAL TESTS

CORRELATION WITH SPECIFIC RESISTANCE

During the latter part of the study, bench-scale filterability tests were performed in conjunction with experimental work on the pilot filter presses. Capillary suction time (CST), modified Buchner funnel (Rv), and high pressure (Rp) methods were used to determine the average specific resistance to filtration of the conditioned sludge mixture and provide correlations with press performance. Detailed descriptions of these test methods are given in Appendix C.

Samples of the conditioned mixture were taken directly from the NGK mix tank to insure that the bench-scale tests were made on the same sludge mixture that was fed to the pilot filters. The filterability tests were begun simultaneously with the start of the press cycles. The CST was measured first, followed by the pressure and modified Buchner funnel determinations, respectively.

In Figures 31-33, the results of these tests for the NGK press are plotted. These graphs show that a definite empirical correlation existed between the average specific resistance of the conditioned sludge mixture and press performance. In general, press yields decreased as the resistance to filtration increased and minimum acceptable filtration for this press, i.e. a process yield of 3.17 $kg/hr/m^2$ to give cake solids of 35%, occurred at $Rv = 27 \times 10^{10}$ cm/g, Rp = 7 and CST = 15 seconds (these values were obtained using a least squares linear regression analysis of the data in Figures 31-33). Results similar to this were also found for each of the fixed volume presses.

At a plant similar to Blue Plains where sludge characteristics and, hence, chemical conditioning demand varies daily, correlations such as these can provide an invaluable tool for controlling full-scale press operations. Tests on the pilot press units showed that the quantitative measure of the specific resistance gave a good indication of press performance, regardless of either sludge blend ratio or quantity of conditioning chemicals added. In the calculation of the specific resistance parameter (See Appendix C.), the effect of these variables is minimized so that consistent values will result. Notice in the following table that comparable resistance values have comparable press yields. For example, the run on the 4/1 secondary/primary sludge required a conditioning dosage of

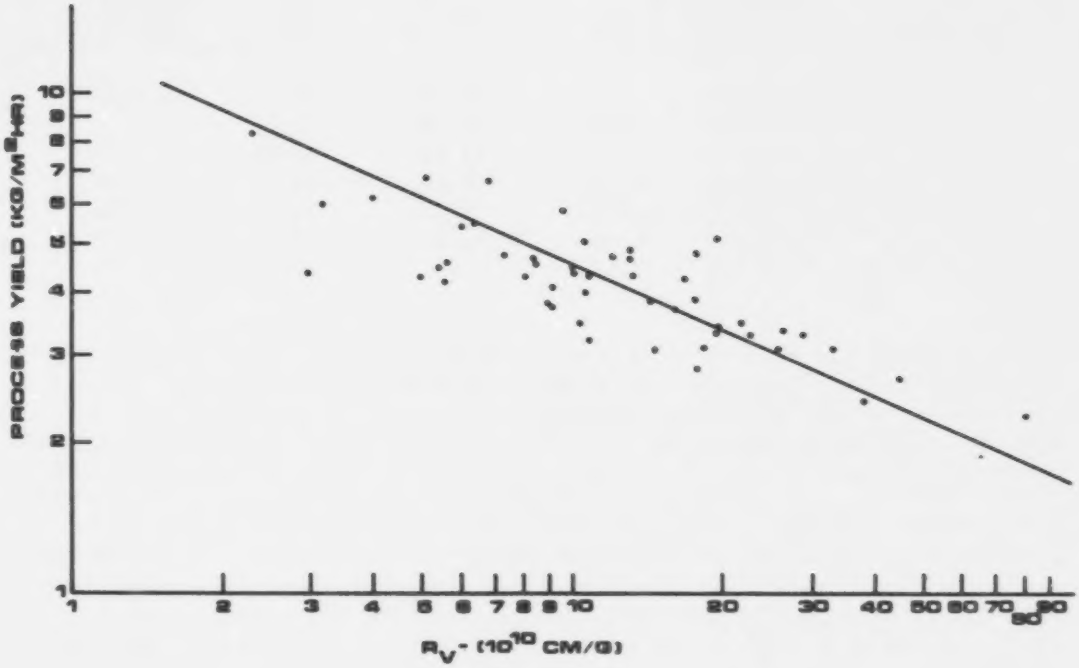


Figure 31. Process yield vs Rv.

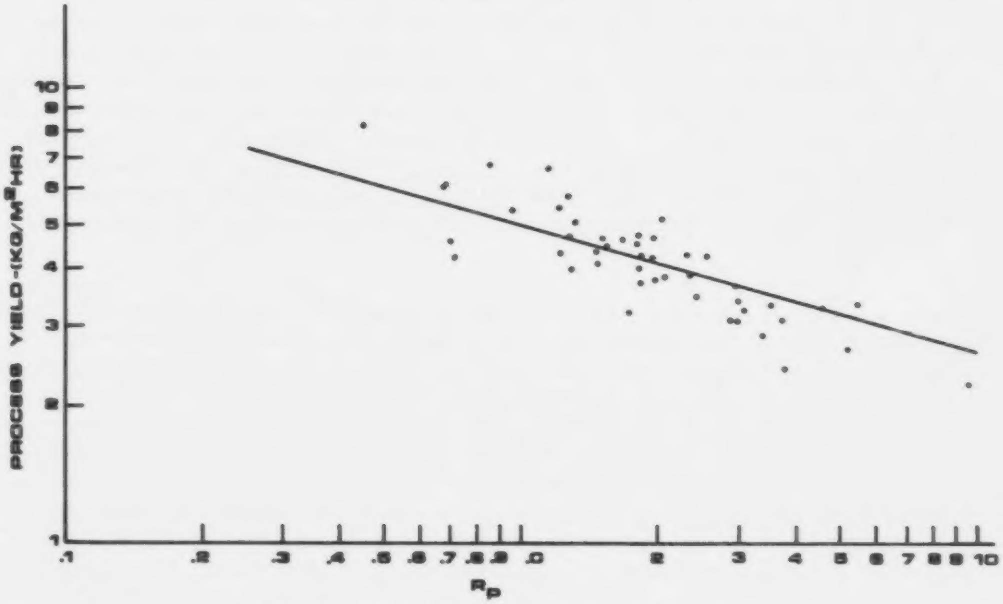


Figure 32. Process yield vs Rp.

RATIO SEC/PRIM	% CHEMICALS Lime/$FeCl_3$	Rp	Rv 10^{10} cm/g	CST sec	PROCESS YIELD kg/hr/m^2
2/1	18.3/6.2	3.02	32.76	12.4	3.12
2/1	16.6/5.5	3.10	28.71	12.5	3.27
2/1	14.2/4.8	-	25.86	15.3	3.12
4/1	29.6/9.9	1.33	10.74	11.0	5.42
1/1	15.2/5.0	1.28	9.57	12.8	5.81
1/2	15.5/5.2	0.86	5.20	9.3	6.79
1/1	18.9/6.4	0.69	4.04	8.4	6.20

29.6% lime and 9.9% $FeCl_3$; specific resistance values of Rp = 1.33, Rv = 10.74 x 10^{10} cm/g, and CST = 11.0 sec were obtained; and a yield of 5.42 kg/hr/m^2 resulted. In the subsequent run on the 1/1 sludge mixture, however, a chemical dosage of only 15.2% lime and 5.0% $FeCl_3$ was required, yet nearly identical resistance values of Rp, Rv, and CST equal to 1.28, 9.57 x 10^{10} cm/g, and 12.8 sec, respectively, were obtained. The resulting press yield, therefore, was also nearly identical at 5.81 kg/hr/m^2. Because of correlations of this type between the specific resistance and the press yield, these bench-scale resistance tests provide a quick method whereby press output can be approximated prior to filtration. And within the span of only a few moments for testing, hours which would be wasted on inadequate filtration can be avoided.

It is evident from the above table, however, that the CST, although a good indicator of filterability, does not give consistent results. Other researchers[2] have found that the test is extremely sensitive to the feed solids concentration of the sludge and, hence, is most useful only when correlated with results from the pressure and modified Buchner funnel tests. In Figures 34 and 35, these correlations, in which the CST has been corrected for feed solids, are shown. While the CST would be the preferred method of determining filterability since it requires only a few seconds to perform, the considerable amount of data scattering suggests that this method introduces a significant error in resistance determinations.

Several manufacturers have indicated that for high pressure filtrations, the pressure method is preferred. In Figure 36, however, the correlation between Rp and Rv shows that the modified Buchner funnel and pressure tests produced comparable results (correlation coefficient = 0.902) for our particular sludge. This indicates that the Buchner test, which requires less time and is much easier to operate, can at times be used with equal accuracy in high-pressure filtration work. Moreover, where precise determinations of actual resistance values are not required, the modified Buchner test can be reduced to a simpler form in which only the quantity of filtrate collected within a given period of time is noted. Compilation

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[2] Baskerville, R.C. and R.S. Gale, "A Simple Automatic Instrument for Determining the Filterability of Sewage Sludges, "Water Pollution Control, 67, 233 (1968).

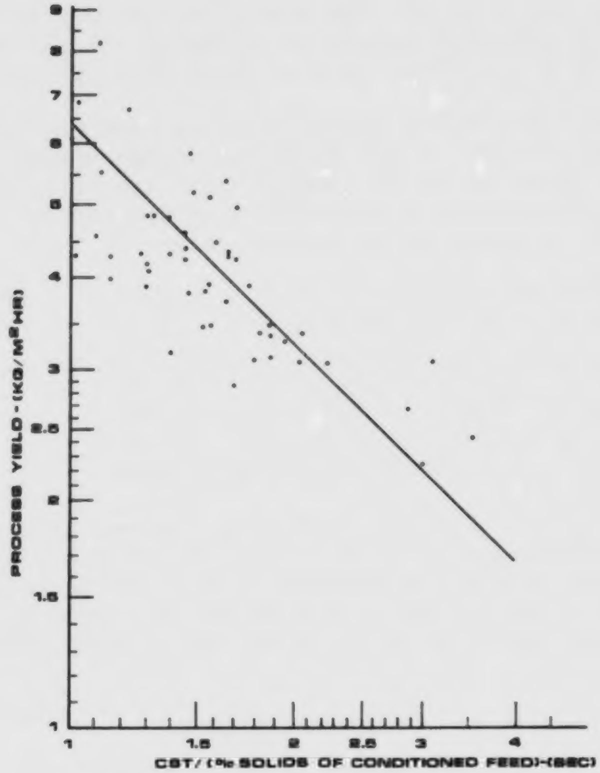

Figure 33. Process yield vs CST/(percent solids of conditioned feed).

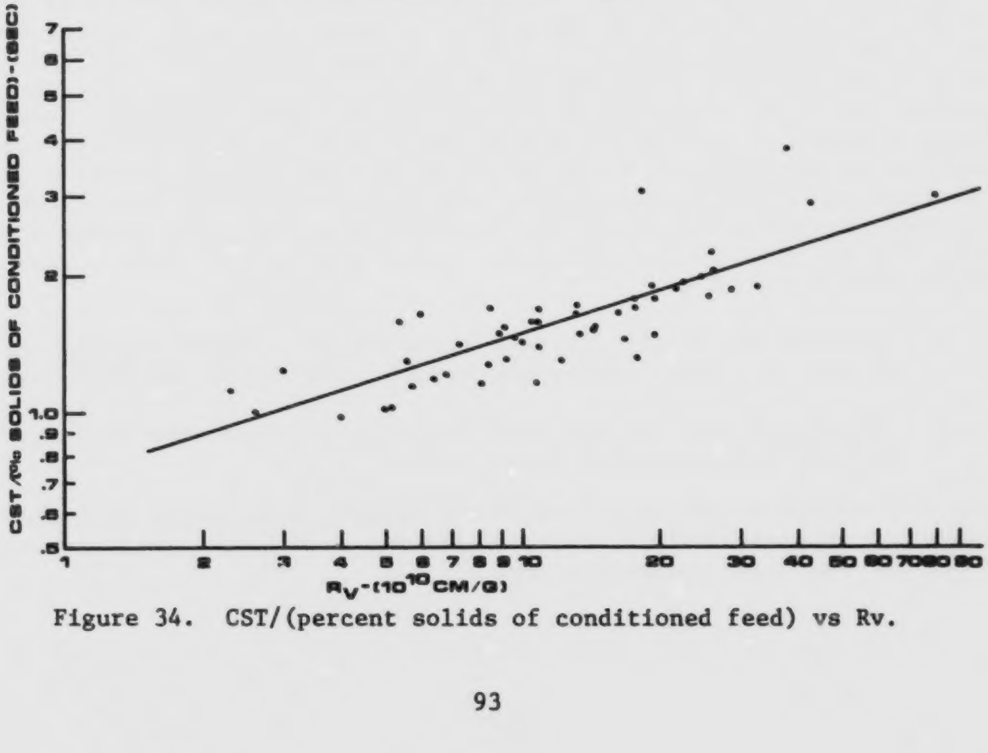

Figure 34. CST/(percent solids of conditioned feed) vs Rv.

of all the data collected for the Blue Plains sludge showed that if 80 mls of filtrate were collected within two minutes, the filter cake would completely form and a resistance value of $Rv = 27 \times 10^{10}$ cm/g would result.

The specific resistance tests were also used to evaluate the effect of varying chemical dosages on the different sludge mixtures. As stated previously and shown in the following table, the filterability of the conditioned sludge generally increased with increasing chemical addition; this change was reflected in decreasing values of the specific resistance:

DATE	% CHEMICALS LIME/$FeCl_3$	Rp	Rv 10^{10} cm/g	CST sec	PROCESS YIELD kg/m^2/hr
8-19	19.6/6.5	1.16	6.84	10.5	6.69
8-10	15.2/5.0	1.28	9.57	12.8	5.8.
8-10	12.1/4.0	2.06	19.49	12.6	5.17

(Ratio secondary/primary = 1/1)

Theoretically, the point at which optimum chemical conditioning occurs, i.e. the greatest increase in press yield per unit addition of chemicals, can be obtained from resistance measurements. In a full-scale installation, with a determination of this type, a substantial cost savings in chemicals can be realized since the unnecessary addition of conditioners would be avoided.

DEWATERING OF VARIABLE SLUDGE CONCENTRATIONS

Throughout the study, the unconditioned sludges averaged 5% total solids. Conditioning with lime and ferric chloride raised the total solids to 6.0% - 6.5% for feeding to the press. Because gravity thickening and air flotation thickening will produce a consistent 5% solids feed, no special tests were run to determine quantitatively the effect that variable feed concentrations had on filter press results; however, the NGK press was capable of handling a range of feed solids from a low of 2.4% (1.8% before conditioning) to a high of 10.0% (8.4% before conditioning). In the low solids region, the press yields were slightly lower because more water had to be processed; but because of the separate squeezing cycle in the diaphragm press, cake solids were not affected. In the high solids region, two adverse effects were noted:

1. Conditioning in the mix tank was difficult because the high sludge solids were very viscous and chemical dispersion was hindered.

2. The sludge pump and the feed ports in the press were more easily plugged with trash and heavy solids.

Total feed solids of 10% appear to be the upper concentration that the diaphragm press can handle. Tests to evaluate feed concentrations were not run for the fixed volume presses.

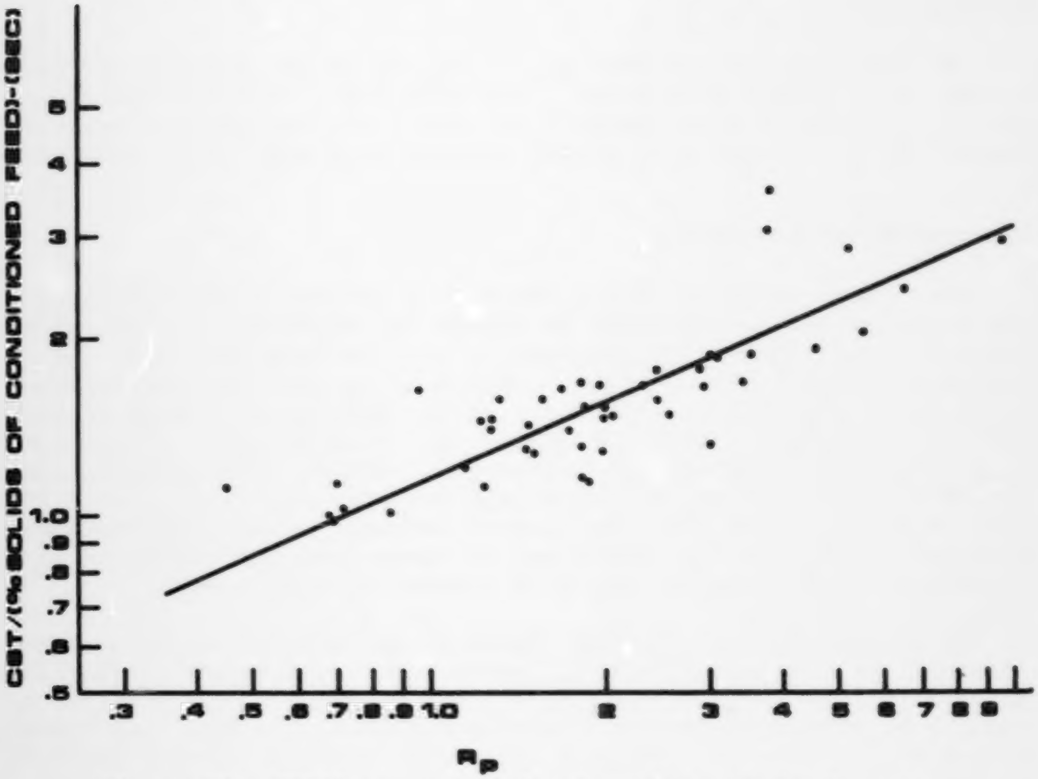

Figure 35. CST/(percent solids of conditioned feed) vs Rp.

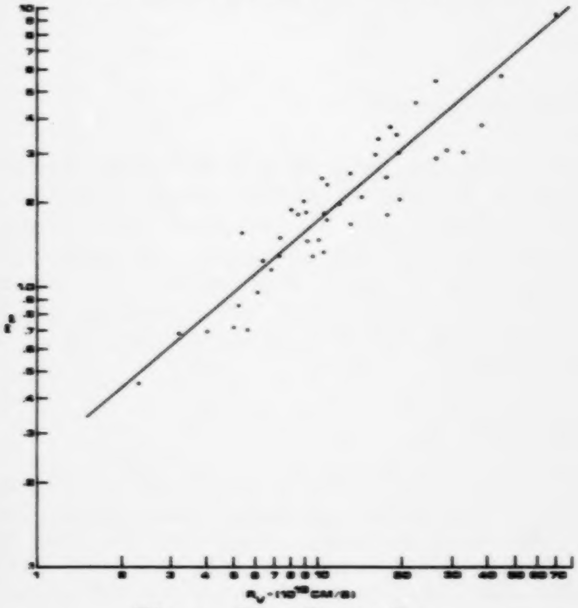

Figure 36. Rp vs Rv.

MATERIAL BALANCE

The NGK press was the only unit tested for which all input and output streams could easily be measured. Test data from a run on 10/18/77 were used to calculate a sample material balance. The calculations, detailed in Appendix D, show a very good balance between input and output total solids.

CONDITIONING WITH POLYMER

Over thirty polymers were screened in an attempt to find a polymer that could adequately condition the sludge for dewatering on a filter press. Allied Colloid's Percol 776 was found to give the best results in the conditioning step. This polymer, a high activity cationic formulation, worked quite well in conjunction with ferric chloride on a vacuum filter. Tests on the Buchner funnel also showed good filterability. Full-scale tests on the NGK press, however, gave rather poor results. Cake solids no greater than 28.6% were achieved, even with extended squeezing cycle times. Yields were, therefore, quite low. The largest problem, however, was the almost immediate cloth blinding. High pressure sprays were ineffective in cleaning the cloths, thus requiring them to be removed for acid washing.

It is believed that the floc formed by the polymer was too weak to withstand the high pressures in the press. But, with proper cloth selection and care in conditioned sludge handling and feeding, the filter press can probably be adapted to dewater a polymer conditioned sludge. The sludge at Blue Plains, however, varies to the extent that one polymer that will work effectively 100% of the time has not yet been found. In contrast, the lime/$FeCl_3$ conditioning system can be adjusted to always give satisfactory results.

TESTS ON PRESS CAKE PROCESSING

The automatic operating mode on the NGK press allowed the production of relatively large quantities of filter press cake for other purposes. Throughout the entire study, the filter press cakes were used for composting trials at the Beltsville, Md. compost site. A number of cakes were analyzed for their calorific value and used in incineration tests conducted in both a multiple hearth incinerator and a rocking grate solid waste incinerator. A local power utility also analyzed the press cake for possible use in coal fired boilers.

Cake Physical Properties

The cake, when discharged from a diaphragm press, resembles a large waffle. Generally, it is rigid and free-standing but breaks up easily (see Figure 37). The density ranges from 1121 to 1185 kg/m^3 (70-74 lb/ft^3 This press cake, when conditioned with lime and $FeCl_3$, dries out in several days. Sludge cakes that were conditioned with 20% lime/6.7% $FeCl_3$ were exposed to ambient weather conditions. One cake was placed in the open, exposed to sunlight, rain, etc. Another cake was placed in the center of a

61 cm (2 foot) high "pyramid" of sludge cakes. Each day a portion of the cake was analyzed for percent solids.

DAY	WEATHER	TEMPERATURE WHEN SAMPLED	% SOLIDS OPEN	% SOLIDS PYRAMID
0	Cloudy	27 °C	39	41.6
1	Rain/clearing	14 °C	44.7	41.1
3	Sunny	20 °C	72	40.8
4	Cloudy	24 °C	77.5	40.7
5	Rain/clearing	24 °C	77.2	41.7

These results show that the press cake does air dry when spread in a thin layer or if stacked vertically. As the cake dried it became impervious to rain and was very hard and brittle. The material in the center of the pyramid did not dry; however, further observations of the cake in the interior of taller piles up to 122 cm (4 feet) showed some self heating after 3 to 4 weeks as aerobic decomposition (composting) proceeded. However, when the cake was broken up into 5 cm (2 inch) pieces prior to piling outside, it was easily rewetted by rain and became difficult to handle. These observations were quite useful when conducting the composting trials.

Cake Breaking

In a large-scale installation, some type of cake breaker will be required prior to any further processing. Fortunately, the diaphragm press gives a fairly uniform product which can be easily handled in a controlled situation. Test work centered around finding acceptable methods of cake breaking and establishing the parameters that affected this step.

Three types of units were found to work:

1. A small tree and branch chipper, operated at high speeds, was capable of breaking up fresh press cakes. The high speeds, however, caused the machine to gum up easily. The unit also reduced any partially dried cake to dust.

2. A garden rototiller, run through the cakes while piled on the ground, was used to prepare the sludge prior to composting. This slow speed unit did an acceptable job for small test quantities of cake.

3. A make shift variable speed screw, pictured in Figure 38, worked quite well to produce chunks in the 5 cm (2 inch) range. Tests on this unit showed that slow speeds gave the best results. The effectiveness of this machine was found to be a function of percent cake solids. Cake solids below approximately 27% tended to stick

Figure 37. Cake from NGK Diaphragm Press.

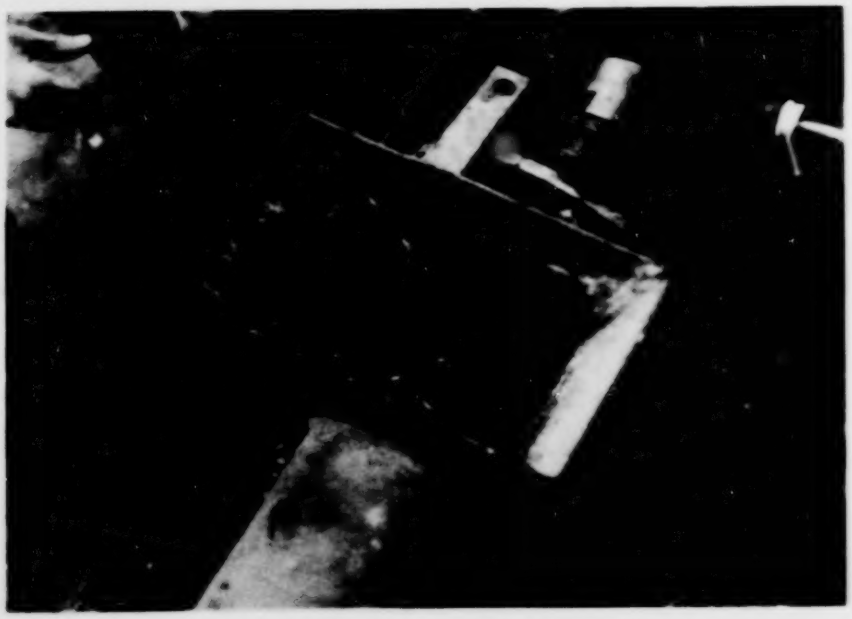

Figure 38. Cake Breaker.

to moving parts. Above the 27% level, the cakes responded to mechanical handling with no sticking. This unit was also effective in breaking up the cakes from the fixed volume presses. The wetter inner sections of the cakes from these presses caused no problems as long as there were sufficient guantities of the dry outside cake sections to help scour the internal screw.

Compost Trials

Tests were conducted at U.S.D.A.'s compost research facility at Beltsville, Maryland, in which the filter press cakes were composted via the static pile method. This method is quite successful for composting vacuum filter cake at 20% solids. For the vacuum filter cake, the sludge is mixed with wood chips (2:1 chips to sludge volumetric ratio; 1:1 weight basis) and stacked to a height of approximately 2.4 m (8 feet) over a perforated pipe. A layer of finished compost blankets the pile. Air is drawn through the pile for a period of 21 days, causing temperatures to reach a normal 70 °C. The mixture is sufficiently deodorized in this time period and the pile is then moved to a stationary curing pile approximately 4.6 m (15 ft) tall for 30 days. The curing period ensures maximum pathogen kill. After screening out the wood chips for reuse, the product is ready for distribution.

The wood chips are needed to reduce the initial moisture of the mix, to provide air passages, and to provide an additional carbon source. The wood chips are the major operating cost of this operation. It was hoped that with the filter press cake, the wood chips could be eliminated or substantially reduced in quantity. Initial composting tests on the press cake without wood chips did produce the required temperatures in the pile, but complete deodorization of the mass was not achieved; the larger chunks of cake had crusted over and contained an anaerobic inner core.

Good results with the filter press cake were obtained by breaking up the cake to a size of 7.6 cm (3 inch) or less with a rototiller, and mixing with wood chips to a volumetric ratio of 0.5:1 chips/sludge cake (approximately 0.2:1.0 weight basis). The same time periods of about 21 days composting and 30 days curing were required for the process. These results, though, are only preliminary since much larger quantities of press cake are required for a full-scale demonstration test. It is projected, however, that if filter press cake is available, cost savings of up to 60% over the vacuum filter operation can be obtained.

Incinerator Tests

A number of the press cake samples were analyzed in a Parr adiabatic oxygen bomb for their calorific value. An average of 32 samples of 2/1 secondary/primary sludge press cake showed that the press cake can be considered to be a low-value fuel that will burn without auxiliary fuel oil.

Dry solids basis	3228 cal/gm (5806 Btu/lb)
Wet solids basis	1225 cal/gm (2204 Btu/lb)
Dry volatile solids basis	6293 cal/gm (11318 Btu/lb)

Multiple hearth unit--

Samples of the filter press cake were incinerated in a 45.7 cm diameter (18 inch) single hearth furnace at the Nichols Engineering Research Facility. The purpose of the tests was to determine if the high chemical content of the Blue Plains filter-pressed sludge would cause any clinkering problems. Prior to one of the tests, the Blue Plains secondary plant was overdosed with $FeCl_3$ to simulate the approximate iron and phosphate content in the sludge that is expected when the advanced waste treatment facilities are completed. This sludge was then overdosed with lime and $FeCl_3$ for conditioning prior to filter pressing. A second sludge tested had the normal amounts of iron and phosphate that were available from the plant at that time. The following table identifies the sludges tested.

SLUDGE CAKE	% CHEMICALS LIME/$FeCl_3$	% Fe*	% SOLIDS	% VOLATILE SOLIDS	BTU/lb DRY SOLIDS
1	24.4/8.2	8	42.8	45.4	4872
2	15.6/5.1	8	39.3	50.1	5373
3	25.9/8.7	5	39.6	47.2	4995
4	13.9/4.6	5	37.7	53.5	5789

*estimated by calculation

Each of the sludge samples were incinerated to complete burnout at temperatures from 927 °C to 1038 °C (1700 °F to 1900 °F). Particle size fed to the furnace ranged from 2.5 to 7.5 cm (1 to 3 inches). Excellent burnout was achieved with no clinker formation. It was concluded that the filter press cake and the high chemical addition would pose no special problems for the incineration of the Blue Plains sludge.

Solid Waste Incinerator Tests--

A qualitative test was conducted to determine if the filter press cake combined with solid waste would burn in a solid waste incinerator. The test unit was a Flynn & Emrich rocking grate design with underfire and overfire air controls. The furnace, fed by cranes from a storage pit, had an average solids detention time of 45 minutes. Approximately 6600 wet kg (3000 lbs) of press cake at 35% solids were dumped into the furnace along with solid waste. The temperature in the combustion chamber above the furnace dropped from its normal 677 °C (1250 °F) to 593 °C (1100 °F) when the sludge was in the burning zone. Examination of the residue, however, showed no signs of the sludge cake and it was assumed to be completely burned. These tests indicated that the press cake would burn well in a co-disposal scheme with solid waste. Because of the press cake moisture, however, there is a limit to the amount that can be blended.

Calculations show that with a 35% solids cake, approximately 20 to 40% of the wet feed to the incinerator can be sludge cake.

Evaluation of Press Cake in a Coal Fired Boiler

Samples of filter press cake were given to a local utility, Potomac Electric Power Company, for their routine fuel analysis. The purpose was to determine if a filter cake-coal mixture could be fed to a boiler to produce electricity. They analyzed the cake sample for ash, sulfur, moisture, and calorific content:

	As received	Dry basis
Ash	12.7%	33.9%
Sulfur	0.14%	0.37%
Water	62.5%	-
Cal/gm (Btu/lb)	1171 (2108)	3123 (5621)

This analysis caused them to reject the press cake as a fuel. They stated that, "Although there are no chemical reasons that we can see which would preclude the use of this sludge as a fuel, the amount of ash is extremely high and would considerably increase our ash handling problems." Because of possible pluggage problems, "our suppliers of coal mills express concern with the fibrous material in the filter press cake.... The amount of gas flow handled by the induced draft fans would increase because of the high moisture content of the sludge. Considering the additional costs for fan power and ash handling and the additional expense of the added new equipment for handling the sludge, it is doubtful if there is any economic benefit to be gained from the burning of sludge. Further study would be required to confirm this preliminary cost estimate."[3] Sufficient quantities of filter press cake were not available for a full-scale test.

[3] Letter R.C. Ungemach (Pepco) to R.C. McDonell (Montgomery County Council), June 3, 1977.

SECTION 8

PROCESS DESIGN

The purpose of the study was to evaluate the various dewatering devices that are capable of producing an auto-combustible cake and to develop design parameters for those units that actually achieved this goal. For auto-combustion, approximately 35% total solids in cakes containing lime and $FeCl_3$ conditioning and 30% solids in polymer-conditioned cakes are required from the dewatering process. The continuous belt press and each of the filter presses met these requirements.

CONTINUOUS BELT PRESS

A continuous belt press can produce a 30% solids cake with polymer conditioning. While theoretically this is acceptable for auto-combustion, some practical problems negated consideration of this press for use at the Blue Plains plant:

1. The need to rely solely on polymer conditioning is unacceptable. During the testing on the belt press, the variability of the sludge feed was so pronounced that no single polymer was found that properly conditioned the feed at all times. Apparently the high-rate secondary process produces a variable waste sludge that has a highly variable response to polymer conditioning. While the use of lime as a conditioning agent would reduce this variability, scaling and cloth plugging problems normally associated with using lime have made belt press suppliers somewhat reluctant to rely on it for conditioning.

2. The high solids content of the filtrate can cause recycle problems in a plant where effluent suspended solids must be controlled to very low levels. Suspended solids capture in the belt press was estimated at 95% for the 2/1 secondary/primary sludge. However, the poor cake discharge generally experienced with this sludge led us to believe that this figure could be an over-estimation of the actual recovery. When advanced waste treatment facilities are completed at Blue Plains, the wastewater effluent must meet a required 7 mg/l suspended solids and 0.22 mg/l total phosphorus standard. Because of the necessity to recycle all water streams from solids processing, the thickening and dewatering systems must have a high degree of solids capture. The entire plant is self-contained with only two effluent streams: the wastewater discharge

to the river and the sludge. If solids are not removed via the sludge stream, they will be recirculated through the treatment system and eventually discharged to the river. Calculations show that if only a 95% solids recovery was achieved in dewatering, the filtrate stream returned to the head of the plant would raise the suspended solids level in the influent raw wastewater by approximately 15 mg/l. This influx of fine solids, together with the recycle solids from thickening operations (which would raise the influent level another 22 mg/l), would pose an additional burden on the wastewater treatment train. Because of the uncertainties in operating advanced waste treatment with multi-media filters, the authors believe that unless a 98-99% solids capture is achieved in the dewatering operation, thus minimizing recycle fines buildup in the system, the plant may have difficulty in achieving its effluent standards.

For these reasons, the belt press was considered unsuitable for use at Blue Plains to dewater thickened sludge. The unit, though, has many advantages that warrent full investigation at other facilities. In a plant that has a fairly consistent sludge that responds well to polymer conditioning, the press can provide a low capital, low operating cost process for producing an auto-combustible cake.

When used as a retrofit device to a vacuum filter, test work showed that the high-pressure section of the belt press can further dewater the vacuum filter cake to the same final cake solids as a filter press. In this case, lime and $FeCl_3$, rather than polymer, were used in conditioning the thickened sludge. No additional chemicals were used to condition the vacuum filter cake prior to dewatering on the belt press. Thus, if the problems encountered during the demonstration of this process can be overcome, the use of the press as a retrofit unit can be a very cost-effective alternative to the filter press, especially for existing vacuum filter installations. Further test work, however, is needed to evaluate possible feeding and distribution systems. Long-range problems associated with the use of lime and $FeCl_3$ on the add-on device should also be assessed.

FILTER PRESS

Chemical conditioning

The addition of lime and $FeCl_3$ to the sludge is necessary for the operation of the filter press. Throughout the study period the chemical dosages required for good filterability varied with the sludge characteristics. In a full-scale continuous dewatering operation one of the highest priorities should be placed on defining the variables in the wastewater processing train that most affect the sludge characteristics and operating the treatment plant in a manner that will minimize their effect on sludge filterability. Doing this will not only provide a smoother dewatering operation, but will also save many dollars in chemical conditioning costs.

Another high priority should be placed on the conditioning step itself. During the course of the study several important large-scale design considerations evolved:

1. Because of (a) the cost of chemicals and (b) the increase in final disposal costs due to the addition of inert conditioning chemicals, optimization of the dosage is necessary. A method of predicting this optimum dosage was not found; however, bench-scale methods (CST, Buchner funnel, pressure tests) were developed which gave an indication of how well the conditioned sludge would dewater on a filter press. To avoid the necessity of running a large number of these tests, the authors feel that a small pilot-model horizontal vacuum filter could be used as a Buchner funnel to continuously monitor the specific resistance filtration parameter. A small slip stream from the conditioning tank would be fed to this filter and, with the unit running at a constant speed, the time needed to produce a dry cake would provide an indication of the dewaterability of the conditioned sludge. The bench-scale tests had shown that if the conditioned sludge could be filtered down to a good cake within 3 to 4 minutes on the Buchner funnel, that same sludge would also filter easily on the pilot press and a 35% solids cake would be produced. By adjusting the feed rate and belt speed of the horizontal vacuum filter, this correlation could be established for the full-scale press. If the cake dries too quickly, the sludge has been over conditioned and the chemicals can be cut back slightly. If the cake takes too long a time to form, the chemical dosage is insufficient and can be increased. The unit would obviously have to be calibrated in the field under continuous operating conditions. A small unit, costing less than $20,000, could provide the necessary information to control a multi-million dollar filter press installation.

   A horizontal vacuum filter was not obtained in time to be tested in the study. When a unit becomes available, however, tests will be conducted to prove this concept.

2. Because of the wide range of sludge feed rates to the press, better control of the conditioning chemicals could be obtained by conditioning at a constant flow rate. The arrangement used during the continuous run on the NGK press (depicted in Figure 16) is a good example. In addition to the conditioning tank, a small surge tank was used to hold the conditioned sludge for feed to the press. The sludge leaving the conditioner could then be sampled and checked for filterability. The conditioning tank must be designed to provide good mixing without shearing the floc. The surge tank should be sized for a maximum of 30 minutes detention time with only enough agitation to keep the solids in suspension. Both tanks must be designed so that miscellaneous trash and small fibers do not build up on the moving parts. A shredder (mazorator) installed in the sludge feed line to the tanks will keep the trash to a manageable size and avoid plugging in the filter press.

3. Careful handling of the conditioned sludge at all times is a necessity if chemical costs are to be optimized. The filter press feed pump must be of a design to minimize shearing of the sludge

floc as the material is delivered to the press under pressure.

4. The corrosiveness of the $FeCl_3$ solution is an important consideration in selecting materials of contruction for both the filter press system and the final disposal system. Sludge cake that has been conditioned with $FeCl_3$ is generally mildly corrosive.

5. During the design phase, close attention must be given to lime slurry handling. Experience at Blue Plains has shown that scaling will occur not only in the lime slurry lines but also in the sludge feed lines and filtrate lines. Injection of an anti-scalant solution in the lime slurry can assist in alleviating these problems. Careful design of all piping systems including access for periodic clean-out is a necessity.

Filter Press Design

The following section on costs shows that the three types of presses--diaphragm, high-pressure, and low-pressure - can all provide the required cake solids at approximately the same unit costs. Several advantages and disadvantages of each type are not, however, readily apparent from these tables.

The diaphragm-type press generally gave higher cake solids, shorter cycle times, and a more uniform cake than the fixed volume presses. Essentially this improved performance was related to cake thickness. The diaphragm press operates best with a 13 to 19 mm (1/2 to 3/4 inch) cake; the low-pressure press with a 25 to 32 mm (1.0 to 1.25 inch) cake; and the high-pressure press with a 30 to 40 mm (1.18 to 1.57 inch) cake. The low-pressure press (100 psig) generally gave shorter cycle times than the high-pressure press (225 psig), although the final overall yield was greater with the high-pressure unit. Apparently, the higher pressures have the advantage of being able to handle the thicker cakes and, therefore, give higher yields per press cycle.

Cake discharge with the high-pressure press was not, however, completely acceptable. Either the selection of a filter media which would improve the discharge or the use of a precoat would be required for a full-scale installation. Cake discharge was generally very good on the low-pressure press and no precoat would be required. The diaphragm press used low pressures to feed the sludge ($<$100 psig), but squeezed at pressures of 213 psig. Cake discharge was always good when the sludge was well conditioned and no precoat would be required.

An additional advantage of the diaphragm-type press is that it is the only type of press that can successfully dewater marginally conditioned sludges. Therefore, during periods of low sludge production, when extra filter capacity is available, less chemicals could be used to marginally condition the sludge. Longer squeeze times would be needed to achieve the required 35% cake solids. Thus, the squeezing diaphragm can be used to minimize the overall chemical costs.

This increased flexibility of a diaphragm-type press can also be used to give any desired cake solids on normally conditioned sludges (by using extended squeeze times) up to the limit of filterability (approximately 40-45% total solids for a 2/1 sludge). The high pressures in the diaphragm press are developed by the squeezing water pump, a relatively low maintenance item. High pressures in a fixed volume press are developed by the sludge feed pump, generally a higher maintenance item.

The production of sludge cake with a diaphragm-type press gives the process advantages as mentioned above; however, from a mechanical standpoint, this also means more mechanical movement of the press components per ton of sludge filtered as compared to the fixed volume presses. This increased mechanical movement could mean not only higher maintenance costs, but also increased instrumentation to control the cycle. In addition, a diaphragm-type press will be discharging cakes every 30-60 minutes, while the fixed volume presses will discharge thicker cakes every three hours. Unless these discharge operations can be made totally automatic and trouble-free, an operator should be present. Accordingly, in the cost estimates for the NGK press, the automatic shaker was eliminated and was replaced by increased manpower. The discharge operation on the diaphragm-type press that releases all cakes at one time (e.g. Ingersoll-Rand, Lasta design) appears to have some advantages over the single or two cake discharge operation.

Diaphragm and cloth replacement costs represent another disadvantage for the diaphragm-type press. Cloth wear occurs in the initial sludge feed portion of each cycle. Once the cake is formed, filtration occurs through the cake and the cloth essentially sees only relatively clear water. Cloth life for all types of presses is estimated at 3000 cycles. With a fixed volume press this equates to changing a set of cloths each year; with a diaphragm press, every three months. Cloth replacement with some filter designs is quite difficult and long periods of downtime are required. Considerably more field testing is required to find more wear resistant cloths and to define the parameters that extend cloth life.

Based on the manufacturer's recommendation, diaphragm life is estimated at 20,000 cycles. Verification of this cycle life must also be established under full-scale operating conditions.

Design Parameters--

NGK diaphragm press--For the 2/1 secondary/primary sludge ratio, Table 10 in the report shows an average of 95 runs over a seven month period. It is assumed that these results would be representative of a full years operation. On the average, the chemical dose would be 20% lime and 7% $FeCl_3$. A 17-minute pump time, 18-minute squeeze time, and a 19-minute mechanical time would be required per cycle (54-minute total). Cake solids of 35-40% would be attained at an average yield of 2.39 $kg/hr/m^2$ (0.49 $lb/hr/ft^2$). The cloth shaker is not included in the mechanical time; it would add another 6 minutes to the cycle for cake discharge. The cloth wash is assumed to be required every 20 cycles.

When more production is required, increased chemical dosages (another 5 percentage points of lime and 2 of $FeCl_3$) can increase the yield by a factor of 1.2 (based on full-scale yield data derived from Table 5). A diaphragm-type filter press installation can therefore be sized for average sludge production figures, but it has a built-in capability to increase sludge throughput by slightly overdosing the chemicals.

The sludge feed system should include a pump to deliver pressures up to 7 kg/cm^2 (100 psig). A feed pressure recorder would provide a useful indication of whether the rate of pressure rise is too great (indicating pluggage or underconditioned sludge). A sludge flowrate meter, correlated with total solids of the sludge feed, could be used to stop the feed pump at the optimum feed rate. The diaphragm pressurization system should include a squeezing water pump that will deliver variable pressures up to 17.6 kg/cm^2 (250 psig). The filtrate during both the pumping and squeezing cycles should be monitored for flowrate and total flow per cycle. The rate monitor would signal the end of either the pumping or squeezing cycle. The flow totalizer would be used to indicate differences between runs and provide a monitor on cloth pluggage.

Based on the information available, the cloth of choice would be the TR 520 type (described in Table 9). The cloth wash system needs further evaluation. The full system pressure of 70 kg/cm^2 (1000 psig) was never achieved during the study. The operating pressure available was only 24.6 kg/cm^2 (350 psig) and this was not always sufficient to clean the cloths. An acid wash system may be necessary for units that use lime for conditioning. Acid washing is quite effective at removing calcuim carbonate and lime deposits both from the cloth and the filtrate passages on the plates. For large-scale installations, both acid washing and high-pressure spray washing should be available. Acid is used to free the system of lime deposits; high pressure sprays are used to remove imbedded sludge particles from the filter media.

The large number of electrical functions necessary for the diaphragm-type press would best be served by using solid state components which could be programmed to indicate malfunctions in both machine and circuit operations. Some problems were encountered with the limit switches and relays that were not easily located and/or correctable.

The NGK pilot press was provided with a 25 mm (1-inch) filtration chamber. Because the sludge was easily filtered on the fixed volume presses with much thicker cakes, we recommend that a thicker chamber be provided on a full-scale design. Chamber thickness up to 38 mm (1.5 inches) should increase the overall yield substantially. A larger chamber would not compromise the advantages of the diaphragm-type press. With easily filterable sludges, more cake per cycle can be discharged. If, however, the sludge filterability is poor, short pump times could still be used to provide a thin, dry cake. Diaphragm and cloth life must each be evaluated with a thicker chamber because of the increased distance of diaphragm movement. An added benefit of a larger chamber is that the feed opening would be less likely to be plugged with rags and trash.

Lasta Diaphragm Press--Comparison testing showed that the Lasta press would give both equivalent cake solids and somewhat higher yields than the NGK press. An average of seven runs on both presses indicated that the full-scale Lasta yield was 22.6% higher than the NGK yield for the Blue Plains sludge. Because of this limited amount of data, design parameters for this press were developed by scaling results from the NGK press. Using the seven month average on the NGK press and applying the 22.6% factor, the full-scale average yield for the Lasta press is then 2.93 kg/hr/m^2 (0.60 lb/hr/ft^2). The number of Lasta filter press units required for installation, though, will be greater than that for NGK. The largest Lasta press has only 204 m^2 filtration area, whereas the largest NGK press has 500 m^2.

The main advantage of the Lasta design is the shorter mechanical turn-around time (10.5 minutes for their 204 m^2 press), since all chambers discharge at once. Additionally, the cake discharge and cloth washing operations are almost completely automatic and the operator attention required would be minimal.

Because of Lasta's shorter mechanical time and, consequently, higher yield, the optimum pumping cycle (as determined from the solids addition rate) is slightly shorter. This provides for a thinner cake and, therefore, somewhat shortened squeezing times. Sufficient data was not collected to compute average cycle times; however, an estimate would be in the range of 30 to 40 minutes total. This assumes that during cloth washing, accomplished by low-pressure (100 psig) sprays, only 1/4 of the filter cloths will be washed each cycle.

The same type of controls as discussed for the NGK press would also be required for the Lasta design. Because of the shorter cycle times in this press, however, the main disadvantages of a diaphragm-type press, i.e. filter cloth and diaphragm replacement costs, could possibly be even more pronounced with the Lasta-type design.

High-pressure press (Passavant)--The full-scale design for the Passavant press is based on the comparison runs in August, 1977. During that time the results showed that the high-pressure press could process the same quantity of sludge but would require 62.3% more filtration area than the NGK press. Design parameters for this press were also developed by scaling the results from the NGK diaphragm unit. Taking the seven month average of data on the NGK press and applying the 62.3% factor gives a Passavant design yield of 1.51 kg/hr/m^2 (0.31 lb/hr/ft^2) with a 40 mm (1.57 inch) chamber thickness. A mechanical time of 20 minutes is required for their Model 20 press (11,625 ft^2). Using 20% lime and 7% $FeCl_3$ for conditioning, cake solids of 34 to 37% will be produced in an average cycle time of 3-1/3 hours. Increasing this chemical dosage (another 5 percentage points of lime and 2 of $FeCl_3$) should result in an increased yield of approximately 20%. As with the NGK press, a built in capacity exists for handling increased sludge production by increasing the chemical dosage.

The sludge feed system should include a pump to deliver pressures up to 15.8 kg/cm^2 (225 psig). A feed pressure recorder is necessary but a flowrate indicator would not be needed with this press. Filtrate rate and total filtrate flow is the preferred method of monitoring the operation. The sludge feed system surge tank should be shared with several presses so that the conditioned sludge detention time does not increase above the 30 minute limit (to avoid floc deterioration).

Low-pressure press (Nichols)--The full-scale design for the Nichols press is based on the comparison runs in August, 1977. During that time the results showed that the low-pressure press could also process an equivalent quantity of sludge but would require 126.8% more filtration area than the NGK press. Again, design parameters were developed by scaling NGK press results because of the limited amount of data available. Taking the seven month data on the NGK press and applying the 126.8% factor gives a Nichols design yield of 1.07 kg/hr/m^2 (0.22 lb/hr/ft^2), with a 32 mm (1.25 inch) chamber thickness. A mechanical time of 20 minutes is required for their largest press (6760 ft^2 with 115 chambers). Using 20% lime and 7% $FeCl_3$ for conditioning, cake solids of 34 to 37% will be produced in an average cycle time of 3 hours. Increasing the chemical dosage will give an increase in yield similar to the high-pressure press. With the exception of using a feed pump of 7.0 kg/cm^2 (100 psig), all other comments for process control are identical to those made for the high-pressure press.

MULTIPLE-HEARTH INCINERATOR DESIGN

Tests run on the single-hearth incinerator showed that there would be no clinkering problem with the filter press cake. We were unable to run any tests to prove that the press cake was auto-combustible; however, the calculations are fairly well known and have been verified in large-scale installations. Figure 39 is a plot showing incinerator outlet temperature as a function of cake total solids and percent conditioners. Any combination that gives an outlet temperature above 800 °F is auto-combustible. The graph was provided by Whitman, Requardt and Associates, Baltimore, Maryland. Assumptions were:

Feed rate: 507,000 lbs/day sludge solids
Volatile solids before conditioning: 60%
Heating value: 10,000 Btu/lb V.S.
Excess air: 75%

The figure shows that with 27% conditioners (20% lime, 7% $FeCl_3$) and 35% cake solids the feed is auto-combustibile with an 800 °F outlet temperature. If the quantity of conditioners increases, then there must be a corresponding increase in percent cake solids. For example, if the dosage rate is increased to 33% (25% lime, 8% $FeCl_3$), then the cake solids must increase to approximately 36.2%. This increase in cake solids is easily accomplished with any of the filter presses. Essentially the figure shows that large increases in chemical addition rates require only small increases in final cake solids to maintain the 800 °F temperature. For a given chemical conditioning, increasing the cake dryness by extending cycle times in the press has the effect of raising the incinerator outlet temperature. If an afterburner is used immediately downstream of the incinerator, this increase in outlet temperature may result in some fuel savings. In

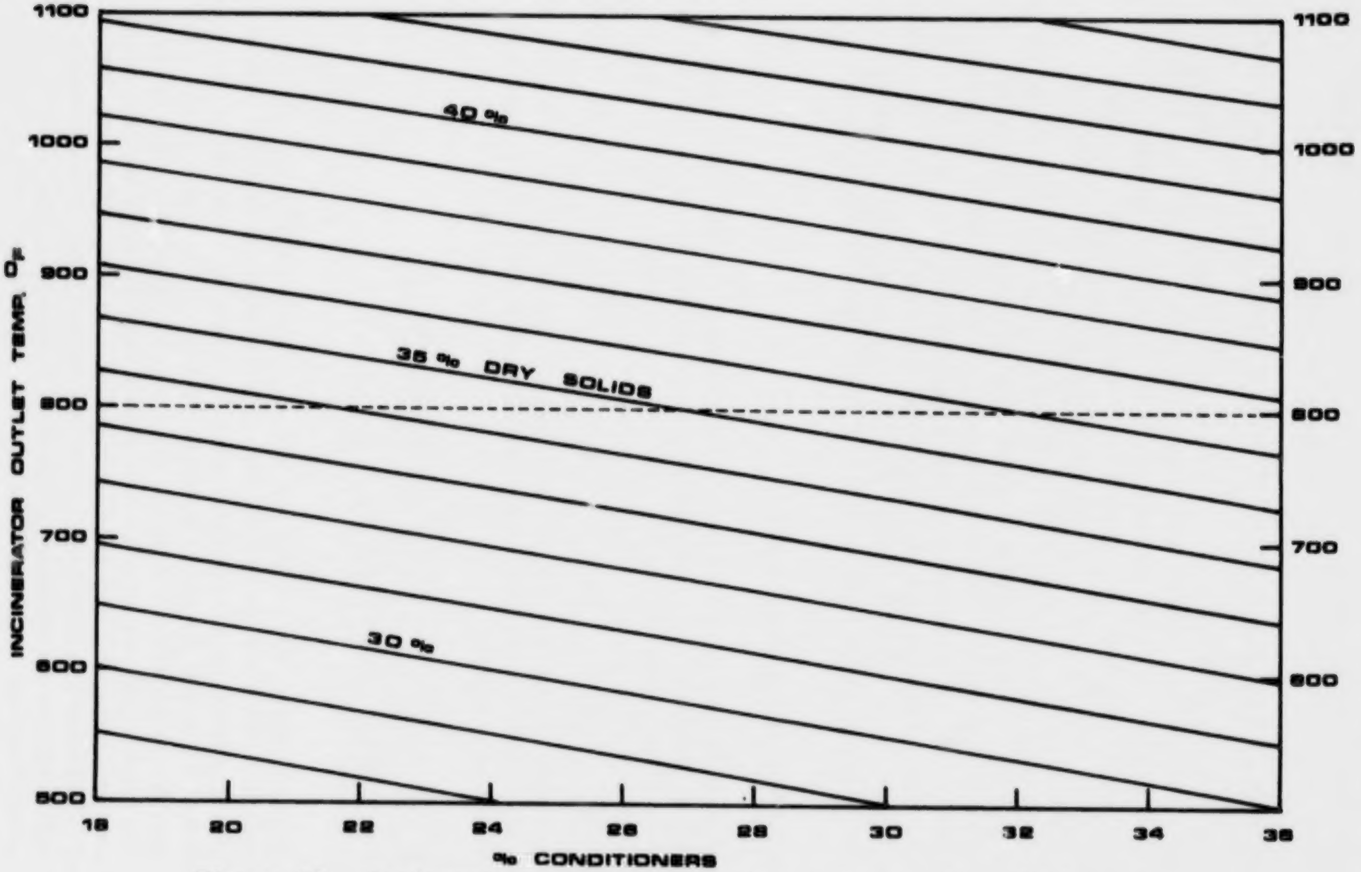

Figure 39. Incinerator outlet temperature vs percent conditioners.

actuality, however, it is more cost-effective to remove the water vapor thus reducing the amount of gas to be heated prior to raising the outlet gas temperature (for toxic pollutant control @ 1350 °F). There is then, little benefit to achieving a very dry cake above the auto-combustible range. However, if a waste-heat boiler is used for steam generation, the higher outlet temperatures could increase the steam production, and in this case there may be some benefit to increasing the cake solids from the press. The alternatives are too complex for any generalized calculations to show the tradeoffs; hence each design must be evaluated on an individual basis.

SECTION 9

DEWATERING AND DISPOSAL COSTS

Estimates of capital and operating costs are presented in Tables 25 through 29. These estimates are for a large municipal wastewater treatment plant generating 250 dry tons of sludge per day (roughly equivalent to a wastewater flow of 200-250 MGD). These estimates are purposely generalized and not specific to the Blue Plains plant. The dewatering options costed are vacuum filters, filter presses, and belt presses. Final disposal costs for both incineration and composting are included.

The following general assumptions were used in developing the tables. The reader is referred to Appendix E for details of all calculations.

1. Sludge: 500,000 lbs/day dry incoming sludge solids @ a concentration of 5% (before conditioning); 2/1 secondary/primary sludge solids ratio.

2. Chemical conditioners: For vacuum filter and filter presses, lime @ 20%, $FeCl_3$ @ 7% of dry sludge solids. Lime cost @ \$0.022/lb; $FeCl_3$ cost @ \$0.065/lb. Anti-scalant needed to help prevent lime deposits. For belt press, polymer costs @ \$15.00 per ton of sludge solids.

3. Yield: Based on test data and expressed as pounds of sludge solids per hour per square foot of filtration area. On belt press, expressed as pounds per hour per meter of belt width.

4. Number of Units: Based on largest size unit available. (See Appendix F for specifications.)

5. Capital cost (1978 dollars): Includes chemical feed system, sludge feed pumps, dewatering unit with all necessary accessories, and conveyor system to transport cake to next process. The total capital cost was obtained by multiplying the manufacturers' equipment cost by a factor of 3 to include installation, piping, utilities, building and engineering.

6. Amortization: Computed at 6-3/8% and 20-year life.
   Capital cost x 0.09 = annual amortization cost.

7. Power: Cost at \$0.04 per kwhr.

TABLE 25. DEWATERING COSTS

	Vacuum Filter	Diaphragm Press	High-Pressure Press	Low-Pressure Press
Sludge Solids (tons/day)	250	250	250	250
Yield (lb/hr/ft^2)	3.0	0.49	0.31	0.22
% Cake Solids	20	35	35	35
Unit Size, ft^2	600	5380	11,625	6760
No. of Units*	13	9	7	15
Capital Cost, $1,000	8,700	23,000	25,350	21,800
Annual Costs, $1,000				
Amortization	783.0	2,070.0	2,281.5	1,962.0
Chemicals	1,663.4	1,663.4	1,663.4	1,663.4
Power	351.9	248.9	210.0	324.5
Water	94.6	6.4	1.6	1.0
Operating Labor	588.0	504.0	420.0	756.0
Maintenance	87.1	496.5	450.0	218.0
Total	3,568.0	4,989.2	5,026.5	4,924.9
Unit Cost, $/ton	39.10	54.68	55.08	53.97

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* Includes one standby unit.

TABLE 26. BELT PRESS COSTS

	Belt Press	Vac. Filter + Belt Press
Sludge Solids (Tons/day)	250	250
Yield (lb/hr/meter of width)	675	3.0 - 1180**
% Cake Solids	30	20 - 35 **
Unit size	3 meter	600 ft^2 - 2 meter **
No. of units*	12	13 - 10**
Capital Cost, $	7,050,000	12,400,000
Annual Costs, $		
Amortization	634,500	1,116,000
Chemicals	1,368,800	1,663,400
Power	81,300	398,240
Water	108,400	153,700
Operating Labor	672,000	840,000
Maintenance	90,500	143,750
Total	2,955,500	4,315,090
Unit Costs, $/ton	32.39	47.29

* Includes one standby unit.
** First entry for vacuum filter; second for belt press

TABLE 27. INCINERATION COSTS

	Vacuum Filter Feed	Filter Press Feed
Total Feed, tons/day	317.5	317.5
% Feed Solids	20	35
% Volatile Solids	47.2	47.2
Furnace Diameter	25' - 9"	25' - 9"
No. Hearths	12	12
Furnace Capacity, lbs wet feed/hr/ft^2	10	10
Capital Cost, $1,000	20,000	10,000
Annual Costs, $1,000		
Amortization	1,800.0	900.0
Power	899.2	405.2
Fuel	4,110.0	630.0
Operating Labor	504.0	336.0
Maintenance	400.0	200.0
Ash Disposal	610.0	610.0
Total	8,323.2	3,081.2
Unit Cost, $/ton of sludge solids	91.21	33.77

TABLE 28. LAND DISPOSAL COSTS

	Vac. Filter Cake @ 20 % solids		Filter Press Cake @ 35% solids	
	$/wet ton	$/dry ton of sludge solids	$/wet ton	$/dry ton of sludge solids
Hauling (25 mile distance)	9.40	59.69	6.25	22.68
Composting	8.84	56.13	6.79	24.64
Total	18.24	115.82	13.04	47.32

TABLE 29. TOTAL DISPOSAL COSTS $/ton of dry sludge solids

	Vacuum Filter Feed	Filter Press Feed
Incineration	130.31	88.45
Composting	154.92	102.00

8 Water: City water, at a cost of $0.53 per 1000 gallons, required for high-pressure washes on the filter presses. Filtered and disinfected water for low-pressure sprays such as required on a vacuum filter and belt press is supplied at a cost of $0.25 per 1000 gallons. Chemical makeup water supplied at no cost from filtrate or plant effluent.

9. Operating labor: $21,000 per man year; 4 crews required per week to cover a 7-day operation. Includes supervision.

10. Maintenance: Based on a percentage of equipment purchase costs plus cloth replacement costs.

11. Unit cost: $/ton of incoming dry sludge solids.

In Table 25, costs for the vacuum filter and the high-pressure and low-pressure press are based on the costs for actual operating installations in the United States and are considered fairly accurate. No large scale operations of the diaphragm-type press are currently on line in the United States; hence, the costs for these units are based on the best information available from the manufacturer. Operating costs for the three types of filter presses are essentially equal at $54 to $55 per ton. Selection of one type versus the other can, therefore, be based on the operating parameters desired and/or competitive bidding. For dewatering only, the vacuum filter provides a cheaper alternative than filter presses. However, the major differences are essentially due to the amortization costs. Out-of-pocket annual operating costs, exclusive of amortization, for either the vacuum filter or any of the filter press types are approximately $30 to $32 per ton.

Table 26 presents some cost estimates for the belt press, both as a single unit or as a retrofit to a vacuum filter. As with the diaphragm press, no large-scale belt press installations are currently on line in the United States to provide actual cost data. Therefore, the estimates given are based on information available from the manufacturers. It is assumed that a suitable polymer at a reasonable cost can be provided for sludge conditioning. The estimates show that the belt press has the potential for providing a very reasonable alternative ($32.39 per ton) to either a vacuum filter or a filter press. However, because of problems detailed in the previous section of this report, the belt press was not considered suitable for Blue Plains.

The use of a belt press as a retrofit to a vacuum filter installation shows a reasonable cost ($47.29 per ton). This estimate assumes the full price for a new vacuum filter installation; enough information is presented in Appendix E to fully cost this option for a specific existing facility. As detailed in the previous section, however, further work must be initiated to develop a workable system prior to implementing this option.

Table 27 shows approximate costs for a multiple-hearth incineration facility. A single train includes a 12-hearth incinerator, electrostatic precipitator, sub-cooler, and fume furnace. Emissions are controlled to the EPA limit of 1.3 pounds of particulate matter per dry ton of solids input. Fuel costs are based on producing an 800 °F exhaust temperature from the furnace and then further raising the stack gases to 1350 °F. For the 20% feed, considerable fuel is required in the furnace; for the 35% feed, the sludge is auto-combustible and fuel is required only to raise the stack gas temperature. Table 27 shows considerable savings when incinerating a cake in the auto-combustible range. Along with the savings in fuel requirements, fewer furnaces (2 vs 4 units) are required, thereby realizing additional savings in power, labor, and maintenance costs.

The hauling costs in Table 28 are based on a 25 mile haul distance to a processing or disposal site. The composting costs are based on the open-air static pile method developed at Beltsville, Maryland. These costs are for processing only, and do not include any costs or revenues derived from the marketing/disposal of the final product. Because of the long transport distance, the costs per dry ton for hauling are nearly equal to the composting costs.

The total disposal costs in Table 29 show:

1. Total disposal costs for filter pressing and incineration are approximately $88 per ton. This compares to the total cost for vacuum filtering and incineration at $130 per ton. Therefore, savings of nearly $4,000,000 per year for a 250 ton-per-day plant are possible by selecting filter presses for dewatering.

2. Total disposal costs for filter pressing and composting (including the cost of hauling the press cake 25 miles) are approximately $102 per ton. This compares to the total cost of vacuum filtering and composting (including hauling) of $155 per ton. Choosing a filter press rather than a vacuum filter, therefore, will result in annual savings of nearly $5,000,000 for a 250 ton-per-day plant.

APPENDIX A

LABORATORY ANALYSES

The following routine laboratory analyses were performed. Detailed descriptions of individual procedures can be found in Standard Methods for the Examination of Water and Wastewater, 14th edition.

Primary, Secondary Sludge	Method
Total Solids	O'Haus moisture balance
pH	Glass electrode
Specific Gravity	Referred to weight of 1 liter of water at room temperature
Sludge Feed	
Total Solids	Dried at 103-105 °C overnight
Volatile Solids	Ignition of dried residue at 610°C
pH	Same as above
Specific Gravity	Same as above
Fe^{+++}	Atomic absorption spectroscopy using Varian AA-6 Spectrophotometer
Filter Cake	
Total Solids	Dried at 103-105 °C overnight
Volatile Solids	Same as above
Density	Variable-volume press - determined from volume of water displaced by a known weight of filter cake Fixed volume press - determined from the total weight of filter cake divided by total chamber volume
Fe^{+++}	Same as above
BTU	Determined using Parr adiabatic Oxygen Bomb Calorimeter
Filtrate	
Total Solids	Same as above

Suspended Solids	Determined according to Standard Methods
pH	Same as above
Cl^-	Determined according to Standard Methods
COD	Determined according to Standard Methods
Total Phosphate	Determined according to Standard Methods
Total Nitrogen, Nitrate	Determined according to Standard Methods

APPENDIX B - DATA SHEETS

DATA SHEET 1

DATE- 6/23/77 PRESS- NGK SQUEEZING PRESSURE- 213 CLOTH TYPE NY51-4

SLUDGE TYPE- Prim & Sec APPEARANCE- Brown color

TEMPERATURE- 25°C GRINDING- Yes PURPOSE OF RUN- Varying chemical conditioning

RUN #	#1	#2	#3	#4	#5
PRIMARY/SECONDARY (pH)$_{pH}$	- / -	- / -	- / -	- / -	- / -
COMBINED SLUDGE-uncond.-cond.	- / -	- / -	- / -	- / -	- / -
LEVEL-before& after cond.(INCH)	$23^{2/3}$ / $21^{1/4}$	23 $^{2/3}$ / 21	$23^{2/3}$ / $21^{3/4}$	/	/
LEVEL (after pumping)	30-7/8	30-1/2	29-1/8		
LIME ADDED (GALS)	16-3/4	20	13-1/4		
FECL3 ADDED(GALS)	5.6	6.7	4-1/2		
SLUDGE PUMPED (GALS)	61.6	60.8	47.2		
FILTRATE COLLECTED (PUMPING)	35.7	39	19.3		
FILTRATE COLLECTED (SQUEEZE)	10.9/46.6	9.9/48.9	11.2/30.5		
FILTRATE COLLECTED (TOTAL)	52.9	53.9	36.5		
FILTRATE:pH	-	-	-		
FILTRATE APPEARANCE	clear	clear	cloudy		
CAKE:WEIGHT (WET)	74	83	64		
CAKE:CONSISTENCY/DISCHARGE	excellent	excellent	wet		
PUMPING TIME	15	16	11		
PUMPING PRESSURE (TERMINAL)	58	53	95		
SQUEEZING TIME	18	18	18		
CLOTH WASH: before run	yes	no	no		
CAKE THICKNESS	1/4-1/2	1/4-1/2	1/4-1/2		
CST (of conditioned sludge)	14.8	10.8	24.2		
CAKE DENSITY	1.1757	1.18	1.12		
TANKED DRAINED	yes	yes	yes		
PRIMARY- SPECIFIC GRAVITY-	1.027				
SECONDARY-SPECIFIC GRAVITY-	1.014				
COMBINED -UNCOND.(SP.GR.)	1.0114	1.0058	1.0027		
COMBINED-COND. (SP.GR.)	1.0208	1.007	1.0065		

PRIMARY-INCHES- 4-1/3 SECONDARY-INCHES- 20 RATIO-#/# PRI./SEC. 32/68

REMARKS: ph meter inoperative

Explanation of Data Sheet 1

1. Information at top of page is self-explanatory. The grinding entry indicates if the primary sludge mazorator was used.

2. pH - was normally measured on each of the sludges, both before and after conditioning, and on the filtrate. The pH probes failed quite often when used with sludge and were inoperative on this particular day.

3. The "LEVEL" entries show the amount of sludge in the tank at various intervals, measured from a reference point at the top of the tank. The total tank depth from this point was 48 inches. Tank calibration was 6.4 gallons/inch.

4. The lime and $FeCl_3$ gallons added are also equivalent to the pounds of each added. Note that lime and $FeCl_3$ used were both 1 lb/gal for easy calibration.

   Lime makeup in the lime slurry tank was as follows:

   An 80 lb bag of lime was added to 75.5 gallons water
   The resultant total volume was 80 gallons.

   $FeCl_3$ makeup in the holding tank was as follows:

   28.6 gallons $FeCl_3$ (30% by weight; 3.5 lbs $FeCl_3$ per
   gallon; Specific gravity = 1.362) was added to the vat.
   The vat was filled to 100 gallons with water and mixed well.

5. The "filtrate collected (pumping)" was measured in gallons by a dipstick reading in the 100 gallon collection vat.

6. The "filtrate collected (squeeze)" was measured in gallons by a dipstick reading in the 15 gallon collection vat. The first entry shows the quantity collected during the squeeze cycle only; the second entry shows the volume collected in the pumping and squeezing cycles combined.

7. The "filtrate collected (total)" shows the total amount in the vats after the pump and squeeze cycles plus the filtrate and sludge blowing cycles. The blow cycles contributed approximately six gallons. The filtrate samples were collected prior to the blowing cycles.

8. Cake weight was measured in pounds by collecting all the discharged cake and weighing on a beam scale.

9. Cake: consistancy/discharge is the operator's opinion on the hardness and quality of the discharged cake.

10. Pumping time is the total time that the sludge pump was running. It takes approximately 1.5 to 2.0 minutes to fill the chambers in the NGK pilot press. Some press manufacturers refer to pump time as the time of filtration after the chambers are filled.

11. Terminal pump pressure is the reading taken on the discharge of the diaphragm pump at the end of the cycle. Because this was a piston pump, the pressure gage pulsed and the reading is only approximate to ± 5 psig.

12. Squeezing time is the total time that the squeezing pump was running. It took only 15-20 seconds to fill the chambers on the pilot press. Larger presses may require 2 to 3 minutes to fill the chambers, thus extending this time in actual large-scale operation.

13. The "cloth wash" is self-explanatory. The automatic system was used.

14. Cake thickness was measured in inches at various points on various cakes.

15. CST (capillary suction time) in seconds shows the relative filterability of the conditioned sludge.

16. Cake density (gm/cc) was measured by placing a one liter graduated cylinder on a balance and measuring both the weight of cake added and the volume of water displaced.

17. "Tank Drained" refers to the status of the mixing tank at the end of the run.

18. The specific gravity of the primary sludge, secondary sludge, unconditioned, and conditioned combined sludges were all measured by weighing one liter of the sample. Because of gas bubbles and pieces of trash in individual samples, this method was not completely accurate. However, the averages of many samples can be used for most calculations with little error.

19. The primary and secondary "inches" refer to volume in the mixing tank. Note the high volume ratio (4.6/1) in contrast to the low weight ratio (2/1).

20. The specific gravity of the 1 lb/gallon chemical solutions were measured periodically.

DATA SHEET 2

DATE: 6 / 23 / 77 Run #1

TIME (MIN)	SLUDGE PUMPING			FILTRATE VOLUME			
				PUMPING		SQUEEZING	
	INCHES	GALLONS	PRESSURE	INCHES	GALLONS	INCHES	GALLONS
0	21 1/4	0	0	0	0	0	0
1	24 5/8	21.6	12	0	0	3 1/4	37.6
2	25 7/8	29.6	21	1 5/8	4.3	5 3/8	38.8
3	27 1/4	38.4	35	3 1/2	9.3	7 5/8	40.1
4	27 3/4	41.6	49	5 3/8	14.3	9 1/2	41.2
5	28 3/8	45.6	49	6 3/8	17	10 7/8	42
6	28 5/8	47.2	50	7 3/8	19.7	12	42.7
7	29	49.6	49	8 1/4	22	13 1/4	43.4
8	29 1/4	51.2	54	9	24	14	43.8
9	29 1/2	52.8	54	9 3/4	26	14 7/8	44.3
10	29 7/8	55.2	54	10 1/2	28	15 1/2	44.7
11	30 1/4	57.6	56	11 1/4	30	16 1/8	45.1
12	30 1/2	59.2	56	11 3/4	31.3	16 5/8	45.3
13	30 5/8	60	58	12 3/8	33	17 1/8	45.6
14	30 3/4	60.8	56	12 7/8	34.3	17 1/2	45.9
15	30 7/8	61.6	58	13 3/8	35.7	17 7/8	46.1
16						18 1/4	46.3
17						18 1/2	46.4
18						18 3/4	46.6
19					BLOWDOWN		52.9
20							↑

COMMENTS:

total cumulative gallons of filtrate

Pump filtrate in 100 gallon vat - 2.67 gal/inch

Squeeze filtrate in 15 gallon vat - 0.58 gal/inch

Explanation of Data Sheet 2

1. During each run, dipstick readings of the sludge and filtrate levels were simultaneously taken by two operators.

2. Column 1 shows the time in minutes.

3. Column 2 shows the inches of sludge in the mixing tank as measured from the top reference point.

4. Column 3 shows the volume of sludge pumped, corresponding to the measurement in column 2.

5. Column 4 shows the sludge feed pump discharge pressure in psig.

6. Column 5 shows the inches of filtrate collected in the 100 gallon vat during the pumping cycle as measured by dipstick from the bottom.

7. Column 6 shows the volume of filtrate corresponding to the readings in column 5.

8. Column 7 shows the inches of filtrate collected in the 15 gallon vat during the squeezing cycle, as measured by dipstick from the bottom. This smaller vat was used so that low filtrate readings could be observed.

9. Column 8 shows the volume of filtrate corresponding to the readings in column 7 and added to the total volume collected during the pumping cycle.

Note that the cycle times for this run were 15 minutes of pumping, followed by 18 minutes of squeezing. Normally, the pumping cycle was terminated when the level in the sludge tank dropped to 1/8" per minute for three consecutive minutes. The squeezing cycle was terminated when the filtrate rate dropped to 1/4" per minute for three consecutive minutes.

DATA SHEET 3

ANALYSIS REQUEST AND REPORT

Analytical Services Laboratory
FWPCA-DC Pilot Plant
Washington, D. C.

Submitted by: W. Ruby

Date Submitted: 6/23/77

Date Reported: 6/24/77

Analysis Requested

LAB NO.	SAMPLE IDENTIFICATION	Unconditioned Feed % Solids	Unconditioned Feed % Vol. Solids	Conditioned Feed % Solids	Conditioned Feed % Vol. Solids	Cake % Solids	Cake % Vol. Solids	Filtrate SS mg/l	Filtrate TS mg/l	
6/23/77	Primary	9.26	71.0							
	Secondary	4.34	60.5							
	NGK R1 Pump	5.15	67.6	6.59	47.5	39.2	47.1	292	9250	
	NGK R1 Squeeze	-	-	-	-	-	-	41	8955	
	NGK R2 Pump	5.68	63.7	7.30	43.8	38.4	44.5	172	10457	
	NGK R2 Squeeze	-	-	-	-	-	-	27	11241	
	NGK R3 Pump	5.90	64.6	6.93	48.9	29.4	47.9	1064	9863	
	NGK R3 Squeeze	-	-	-	-	-	-	422	9001	

Explanation of Data Sheet 3

The following analyses were determined for each filter press run. Laboratory procedures are explained further in Appendix A.

1. Primary sludge - % total solids
   % volatile solids

2. Secondary sludge - % total solids
   % volatile solids

3 3. Combined sludge - unconditioned feed
   % total solids
   % volatile solids

4. Combined sludge - conditioned feed
   % total solids
   % volatile solids

5. Filter Cake - % total solids
   % volatile solids

6. Filtrate - total solids as mg/l
   suspended solids as mg/l

The primary and secondary sludge samples were composites taken either from the NGK mixing tank or directly from the discharge of the thickeners.

The combined sludge samples, both unconditioned and conditioned, were dipped from the NGK mix tank.

Filter cake samples were taken at random from various sections of at least four of the six cakes. Tests were run to show that the cakes were uniform throughout with respect to % solids. No appreciable difference in % solids of the six cakes was ever observed.

Filtrate samples were dipped from the filtrate collection tanks after agitation to ensure a representative sample. For some runs the filtrates from the pumping and squeezing cycles were analyzed separately.

DATA SHEET 4

FILTER PRESS DATA - NGK

DATE- RUN# - 6/23/77	1	2	3	
TYPE SLUDGE-	Prim & Sec →	→	→	
RATIO- #/#-PR./SEC.	31.7/68.3 →	→	→	
FEED SOLIDS-%SOL./%VOL÷(uncond.)	5.15/67.5	5.68/63.7	5.9/64.6	
PRIMARY-%SOL./%VOL/ pH.-	9.26/71/-	9.26/71/-	9.26/71/-	
SECONDARY-%SOL./%VOL./ pH.-	4.34/60.5/-	4.34/60.5/-	4.34/60.5/-	
FEED SOLIDS-%SOL./%VOL.-(cond.)	6.59/47.5	7.30/43.8	6.93/48.9	
PRIMARY- SP. GR. (gr./cc.)	1.027	1.027	1.027	
SECONDARY-SP.GR.-(gr./cc.)	1.014	1.014	1.014	
FEED SOLIDS-SP.GR.-(uncond.)- pH.-	1.0114/-	1.0058/-	1.0027/-	
FEED SOLIDS-SP.GR.-(cond.)- pH.-	1.0208/-	1.007/-	1.0065/-	
LIME (added) %-	24.7	27.0	17.2	
FECL3 (added) %-	8.3	9.0	5.9	
VOLUME-(feed to press)-GALS.	61.6	60.8	47.2	
PUMP TIME-(Mins.)	15	16	11	
SQUEEZE TIME-(Mins.)-	18	18	18	
TERMINAL PRESSURE-Pump psig.-	58	53	95	
SQUEEZING PRESSURE-psig	213	213	213	
FILTRATE VOLUME-(gals.)PUMPING-	35.7	39.0	19.3	
FILTRATE VOLUME-(Gals.) SQUEEZE-	10.9/46.6	9.9/48.9	11.2/30.5	
FILTRATE VOL.-(Gals.) TOTAL+B.D.	52.9	53.9	36.5	
FILTRATE pH.	-	-	-	
FILTRATE (mg/1) TOTAL SOLIDS- pump/squeeze	9250/8955	10457/11241	N.S.	
FILTRATE (mg/1) SUS. SOLIDS- pump/squeeze	292/41	172/27	N.S.	
FILTER CAKE-(Wet weight)	74	83	64	
FILTER CAKE-(%Sol./%Vol.)	39.2/47.1	38.4/44.5	29.4/47.9	
FILTER CAKE (Dry Weight) corr.	21.8	23.4	15.3	
CAKE THICKNESS-(Inches)	1/4-1/2	1/4-1/2	1/4-1/2	
YIELD (lbs./ft.2hr.) process	0.63	0.66	0.50	

* NS indicates no sample was taken

CLOTH TYPE-NY51-4 TEMPERATURE- 25°C PRESS MECHANICAL TIME-5 min. PRESS FILTER AREA- 62.4 ft^2

CONCLUSIONS- Tests show good filtration and high yields with high chemical dosage. Low lime (17.2%) would not dewater effectively. Note low yields and low cake solids.

Explanation of Data Sheet 4

Data Sheet 4 combines the data collected from the run (Data Sheets 1 and 2) with the laboratory results (Data Sheet 3) and several calculations to summarize the series of tests run. All examples use Run #1.

1. The actual weight ratio of primary to secondary sludge was calculated from the measurements taken of the volume, specific gravity, and laboratory % solids of each sludge.

$$\text{lbs solids} = (\text{inches in tank}) \times (\text{tank calibration}) \times (\text{specific gravity}) \times (\text{density of water}) \times \frac{\%\ \text{solids}}{100}$$

$$\text{lbs primary sludge} = 4.33\ \text{in} \times 6.4\ \text{gal/in} \times 1.027 \times 8.3453\ \text{lb/gal} \times \frac{9.26}{100} = 22.0\ \text{lbs}$$

$$\text{lbs secondary sludge} = 20\ \text{in} \times 6.4\ \text{gal/in} \times 1.014 \times 8.3453\ \text{lb/gal} \times \frac{4.34}{100} = 47.0\ \text{lbs}$$

$$\text{Ratio: lbs primary/lbs secondary} = \frac{22}{47} \text{ or } \frac{31.7\%}{68.3\%}$$

2. Percentages of total and volatile solids, pH (if available), and specific gravity were summarized.

3. The percentages of lime and $FeCl_3$ were calculated as follows. In computing the chemical dosages, the total lbs of solids in the tank were determined from the feed solids (unconditioned) measurement and the volume.

$$\text{lbs solids} = 24.33\ \text{in} \times 6.4\ \text{gal/in} \times 1.0114 \times 8.3453\ \text{lb/gal} \times \frac{5.15}{100} = 67.7\ \text{lbs}$$

Lime added = 16.75 gal x 1 lb/gal = 16.75 lbs

$$\%\ \text{lime} = \frac{16.75}{67.7} \times 100 = 24.7\%$$

$FeCl_3$ added = 5.6 gal x 1 lb/gal = 5.6 lbs

$$\%\ FeCl_3 = \frac{5.6}{67.7} \times 100 = 8.3\%$$

4. Filter cake corrected dry weight is given. The final cake weight was

corrected for the chemicals added.

$$\text{Cake dry weight} = \text{wet weight} \times \frac{\%\ \text{solids}}{100}$$

$$= 74\ \text{lbs} \times \frac{39.2}{100}$$

$$= 29\ \text{lbs}$$

Corrected dry weight = (cake dry weight)-(chemical weight)
The chemical weight is known as a percentage of the incoming sludge solids; the proper formula is then:

$$\text{Corrected dry weight} = \frac{\text{cake dry weight}}{1 + \frac{\%\ (\text{lime} + FeCl_3)}{100}}$$

$$= \frac{29}{1 + \frac{(24.7 + 8.3)}{100}}$$

$$= 21.8\ \text{lbs}$$

5. Yield is reported as:

$$\frac{\text{corrected dry weight}}{(\text{cycle time}) \times (\text{filtration area})}$$

Corrected dry weight = 21.8 lbs

$$\text{cycle time} = 15\ \text{min} + 18\ \text{min} = \frac{33\ \text{min}}{60\ \text{min/hr}} = 0.55\ \text{hr}$$

Filtration area = 62.4 ft^2

$$\text{Yield} = \frac{21.8\ \text{lbs}}{(0.55\ \text{hr}) \times (62.4\ ft^2)}$$

$$= 0.63\ \text{lb/hr/}ft^2$$

This yield is called the process yield, since it includes only pump and squeezing times. It is a good figure for comparing runs in a set to establish trends. For scale-up, the mechanical time of the full-scale press must be included in the "full-scale yield". For the 500 m^3 NGK press, this mechanical time is 19 minutes. It includes time to open and close the press, discharge cake, and fill the squeezing chambers, etc. The full-scale cycle time is then:

$$33\ \text{min} + 19\ \text{min} = \frac{52\ \text{min}}{60\ \text{min/hr}} = 0.87\ \text{hr}$$

$$\text{Full-scale yield} = \frac{21.8\ \text{lbs}}{(0.87\ \text{hr}) \times (62.4\ ft^2)} = 0.40\ \text{lbs/hr/}ft^2$$

DATA SHEET 5

PASSAVANT PRESS - OPERATIONAL DATA

DATE 8/25/77

RUN #	1
REFERENCE RUN-(NGK)	#1
CLOTH TYPE	
SLUDGE TYPE	Sec + Prim
RATIO-PR./SEC.	31.6/68.4
TEMPERATURE	28° C
CST (COND. SLUDGE)	-
SLUDGE pH (UNCOND./COND.)	-
LIME %	20.1
FERRIC %	6.7
FILTRATE VOLUME(TOTAL)	39.7 gal
FILTRATE pH.	-
FILTRATE APPEARANCE	-
CAKE (DISCH./CONSIST.)	good
CLOTH COND.(AFTER DISCH.)	some sticking
CAKE WEIGHT	#1 #2
CAKE WEIGHT TOTAL	87.5 lbs
CAKE THICKNESS	1.5 inches
TOTAL CYCLE TIME (MINS.)	160

TIME	PRESSURE	FILTRATE VOLUME	FILTRATE RATE
0	0	inches	0 (gal)
5	110	3-5/8	9.7
10	170	5-1/2	14.7
15	215	6-3/4	18.0
20	225	7-7/8	21.0
25	225	8-5/8	23.0
30	225	9-1/4	24.5
35			
40	225	10-3/8	27.7
45			
50	225	11-1/8	29.7
55			
60	225	11-7/8	31.7
65			
70	225	12-1/2	33.4
75			
80	225	12-3/4	34.0
85			
90	225	13-1/4	35.4
95			
100	225	13-1/2	36.0
105			
110	225	13-7/8	37.0
115			
120	225	14-1/4	38.0
125			
130	225	14-3/8	38.4
135			
140	225	14-5/8	39.0
145			
150	225	14-7/8	39.7
155			
160	225	14-7/8	39.7

REMARKS: Comparison with NGK and Nichols.

Explanation of Data Sheet 5

Data Sheet 5 was used to collect information during each run on the Passavant press.

1. The data in the column on the left is self-explanatory. Some information was taken from the referenced corresponding NGK run.

2. Pressure and filtrate readings were normally taken every ten minutes during the run. Filtrate was collected in a 100 gallon plastic vat. Vat calibration was 2.67 gal/inch.

DATA SHEET 6

PASSAVANT PRESS-DATA SHEET

DATE 8/25/77				
RUN #	1			
TYPE SLUDGE:	Prim/Sec 1/2			
RATIO: # Pr./ Sec.	31.6/68.4			
FEED SOLIDS:%SOL./% VOL.(uncond.)/pH.	5.65/60.4/-			
FEED SOLIDS:%SOL./% VOL.(cond.)/ pH.	7.59/42.9/-			
LIME %:	20.1			
FERRIC %:	6.7			
CYCLE TIME:(MINS.)	160			
TERMINAL FEED PRESSURE(psig):	225			
FILTRATE VOLUME: (gal.)	39.7			
FILTRATE pH.:	-			
FILTRATE:(mg/l) TOTAL SOLIDS	8831			
FILTRATE:(mg/l) SUS.SOLIDS	29			
FILTER CAKE:(WET WEIGHT)	87.5 lbs			
FILTER CAKE:(%SOL./ %VOL.)	36.2/44.2			
FILTER CAKE:(CORR. DRY WEIGHT)	25.0 lbs			
CAKE THICKNESS:(INCHES)	1.5			
CAKE DENSITY: (lbs./ft.^{3})	76.8			
YIELD: (lbs./ft.^{2}/hr.) Process	0.52			
YIELD: (full scale)	0.46			
Terminal filtrate rate (gal/hr/ft^2)	0			

CLOTH TYPE: T-167 TEMPERATURE: 28 °C PRESS AREA: 18.0 ft^2

CONCLUSIONS:

Good run for 2/1 sludge

Explanation of Data Sheet 6

Data Sheet 6 was used to summarize all data collected on each run on the Passavant press.

1. Data on the sludge ratio, the % feed solids, and the chemical rates were taken from the corresponding NGK run.

2. Filtrate solids were determined from a composite sample of the filtrate.

3. Cake solids were determined by averaging the % solids for each of the two cakes sampled.

4. Corrected dry weight was determined by the same procedure described in the explanation for Data Sheet 4.

5. Process yield was reported as:

$$\frac{\text{Corrected dry weight}}{(\text{cycle time}) \times (\text{filtration area})}$$

$$\text{Yield} = \frac{25.0 \text{ lbs}}{(160/60 \text{ hr}) \times (18.0 \text{ ft}^2)}$$

$$= 0.52 \text{ lb/hr/ft}^2$$

6. Full-scale yield was calculated by adding 20 minutes to the process cycle time (turn around time required for a large-scale press).

$$\text{Full-scale cycle time} = 160 \text{ min} + 20 \text{ min} = \frac{180 \text{ min}}{60 \text{ min/hr}} = 3.0 \text{ hr}$$

$$\text{Full scale yield} = \frac{25.0 \text{ lbs}}{(3.0 \text{ hr}) \times (18.0 \text{ ft}^2)}$$

$$= 0.46 \text{ lb/hr/ft}^2$$

DATA SHEET 7

NICHOLS PRESS - OPERATIONAL DATA

DATE 8/25/77	
RUN #	1
REFERENCE RUN-(NGK)	#1
CLOTH TYPE	4709/40
SLUDGE TYPE	Sec + Prim
RATIO-PR./SEC.	31.6/68.4
TEMPERATURE	28° C
CST (COND. SLUDGE)	-
SLUDGE pH (UNCOND./COND.)	-
LIME %	20.1
FERRIC %	6.7
FILTRATE VOLUME(TOTAL)	23,250 ml
FILTRATE pH.	-
FILTRATE APPEARANCE	clear
CAKE (DISCH./CONSIST.)	excellent
CLOTH COND.(AFTER DISCH.)	clean
CAKE WEIGHT grams	#1 2160 #2 2175
CAKE WEIGHT TOTAL	4335 gms - 9.55 lbs
CAKE THICKNESS	1"
TOTAL CYCLE TIME (MINS.)	130

TIME	PRESSURE	FILTRATE VOLUME	FILTRATE RATE
0			
5	100	8100	8100
10	115	12300	4200
15			
20	95	16250	3950
25			
30	110	18600	2350
35			
40	85	20050	1450
45			
50	107	21100	1050
55			
60	93	21650	550
65			
70	100	22090	440
75			
80	103	22430	340
85			
90	95	22690	260
95			
100	101	23890	200
105			
110	107	23030	140
115			
120	100	23150	120
125			
130	101	23250	100
135			
140			
145			
150			
155			
160			
165			
170			
175			
180			

REMARKS: Comparison with NGK and Passavant

Time (min)	Pressure (psig)	Filtrate rate (ml)	Total Volume (ml)
1	15	850	850
2	30	2500	3350
3	50	1500	4850
4	75	1950	6800
5	100	1300	8100

DATA SHEET 8

NICHOLS PRESS-DATA SHEET

DATE 8/25/77				
RUN #	1			
TYPE SLUDGE:	Prim/Sec 1/2			
RATIO: #/# Pr./ Sec.	31.6/68.4			
FEED SOLIDS:%SOL./% VOL.(uncond.)/pH.	5.65/60.4/-			
FEED SOLIDS:%SOL./% VOL.(cond.)/ pH.	7.59/42.9/-			
LIME %:	20.1			
FERRIC %:	6.7			
CYCLE TIME:(MINS.)	130			
TERMINAL FEED PRESSURE(psig):	101			
FILTRATE VOLUME:(MLS.)	23250			
FILTRATE pH.:	-			
FILTRATE:(mg/l) TOTAL SOLIDS	8013			
FILTRATE:(mg/l) SUS.SOLIDS	49			
FILTER CAKE:(WET WEIGHT)	9.55 lbs			
FILTER CAKE:(%SOL./ %VOL.)	36.2/47.3			
FILTER CAKE:(CORR. DRY WEIGHT)	2.7 lbs			
CAKE THICKNESS:(INCHES)	1			
CAKE DENSITY: (lbs./ft.^{3})	71.5			
YIELD: (lbs./ft.^{2}/hr.) process	0.31			
Terminal filtrate rate ml/min	10			
Full scale yield (lbs/ft^2/hr)	0.27			

CLOTH TYPE: 4709/40 TEMPERATURE: 28° C PRESS AREA: 4 ft^2

CONCLUSIONS:

Good run for 2/1 sludge

Explanation of Data Sheet 7

Data Sheet 7 was used to collect information during each run on the Nichols press.

1. This data sheet is nearly identical to Data Sheet 5 and the same remarks are appropriate.
2. The first five minutes of the run were programmed to increase the pressure in the feed tank slowly by using a regulator valve.
3. Pressure and filtrate readings were normally taken every ten minutes during the run. Filtrate was collected in a graduated cylinder and composited during the run. The pressure fluctuated slightly because of other demands on the air supply system.

Explanation of Data Sheet 8

Data Sheet 8 was used to summarize all data collected on each run on the Nichols press. This data sheet is identical to Data Sheet 6.

APPENDIX C

DETERMINATION OF SPECIFIC RESISTANCE TO FILTRATION

From June through October 1977, laboratory measurements of the filterability of the conditioned sludge mixtures were made using three different techniques:

1. Modified Buchner funnel method
2. Positive pressure method
3. Capillary suction method

The modified Buchner funnel and pressure methods give quantitative measurements of the filterability of a sludge by determination of its average specific resistance to filtration. The specific resistance is defined as the pressure difference required to produce a unit rate of filtrate flow of unit viscosity through a unit weight of cake and is calculated from the following equation:

$$R = \frac{2PA^2b}{\mu c}$$

Where R = average specific resistance
P = the filtration pressure
A = the filter area
μ = the viscosity of the filtrate
c = the weight of cake solids per unit volume of filtrate
b = slope of filtrate discharge curve, i.e. time/volume versus volume

A more detailed discussion of the theoretical derivation of this equation can be found in the literature.[4]

The capillary suction test was developed as an alternative to the Buchner funnel and pressure tests. It measures the time required for filtrate to drain from a given volume of sludge and is easily correlated with the specific resistance parameter.

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[4] Carman, P.C. "Fundamental Principles of Industrial Filtration (A Critical Review of Present Knowledge)." Transactions of the Institution of Chemical Engineers, 1938, 16, 168.

MODIFIED BUCHNER FUNNEL METHOD

This technique is used most frequently in studies of sewage sludge filtrations.[5,6] Essentially, it consists of effecting the filtration process by the use of a vacuum pressure.

Facilities

The filter consisted of a white, porcelain Buchner funnel (9.1 cm plate diameter). Standard laboratory paper (9 cm diameter) graded for a fast filtering speed was used as the filter media. Suction was provided by a 24 inch gage vacuum pump. Accessory equipment included a 250 ml graduated cylinder with side arm, stand, stopwatch, and thermometer. The funnel, with accessories, was assembled as shown in Figure C-1.

Operation

A filter paper was wetted and placed in the bottom of the funnel. The conditioned sludge was prepared by pouring twice from one beaker to another to resuspend any solid particles which had settled out. The temperature of the sample was noted and a 200 ml portion was measured into the funnel. The vacuum was then started immediately, and simultaneous readings of filtrate quantity and time were recorded as the filtrate collected. These readings were taken at 10 or 15 second intervals for an elapsed time of 240 seconds or until the vacuum broke, i.e. the filter cake was completely formed.

Analysis

$$R = Rv = \frac{2PA^2b}{\mu c}$$

Experimental parameters were measured in the metric system and Rv reported in units of cm/g. A filtration vacuum differential of 24 inches Hg (81360

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5. Coackley, P. and B.R.S. Jones. "Vacuum Sludge Filtration I. Interpretation of Results by the Concept of Specific Resistance. Sewage and Industrial Wastes, 1956,28, 963.

6. Swanwick, J.D., F.W. Lussignea, and K.J. White. "The Measurement of the Specific Resistance to Filtration and Its Application in Studies of Sludge Dewatering." Journal of the Institute of Sewage Purification, 1961, 6 487.

N/m^2) was used. The filtration area was taken to be that of the filter paper, 63.62 cm^2. The filtrate viscosity, μ, was assumed to be that of water, measured at the temperature of the sample and converted to units of Ns/m^2. The weight of dry cake solids per unit volume of filtrate, c (g/cm^3), was approximated from the sample feed solids concentration according to the relationship:

$$c = \frac{TSS}{1000-TSS}$$

where TSS is the total suspended solids of the feed, g/l.

The slope of the filtrate curve, b (s/cm^6), was obtained by plotting θ/V vs V, where V is the filtrate volume collected in time θ. Carman[7] noted that in the determination of b, the initial readings of θ and V represent the initial resistance of the filter medium rather than the specific resistance of the solids. Hence, θ and V should not be measured from the beginning of the filtration; rather, the filtration pressure should be raised slowly to its full value to minimize the effect of this initial resistance. Once the pressure reaches constancy, the readings should then be taken. In our determinations, this procedure was not used; pressures were raised immediately to full value and measurements of θ and V were taken from the start of the experiment. However, in order to circumvent the problem posed by resistance of the filter medium, these first initial readings were not used in computing the slope of the filtrate curve. By doing this, the procedure was simplified and standardized for all operating personnel, yet experimental accuracy was still maintained.

A sample of the data collected and its analysis is shown for this method in the following data sheet.

POSITIVE PRESSURE METHOD

This method has been used by several researchers[8,9] during studies of high-pressure filtration. It is similar in operation to the modified Buchner funnel method, except that the filtration pressure is provided by positive instead of vacuum pressure.

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[7] Carman, P.C. op. cit.

[8] Coackley, P and B.R.S. Jones. "Vacuum Sludge Filtration I. Interpretation of Results by the Concept of Specific Resistance." Sewage and Industrial Wastes, 1956, 28, 963.

[9] "Pressure Filtration of Waste Water Sludge with Ash Filter Aid." Environmental Protection Agency (EPA) Technology Series, 1978, EPA-R2-73-231.

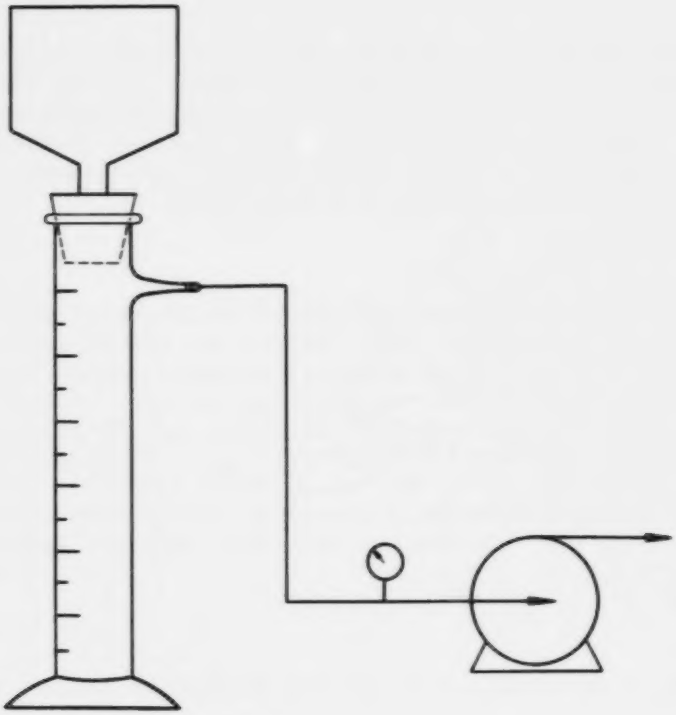

Figure C-1. Buchner Funnel Apparatus.

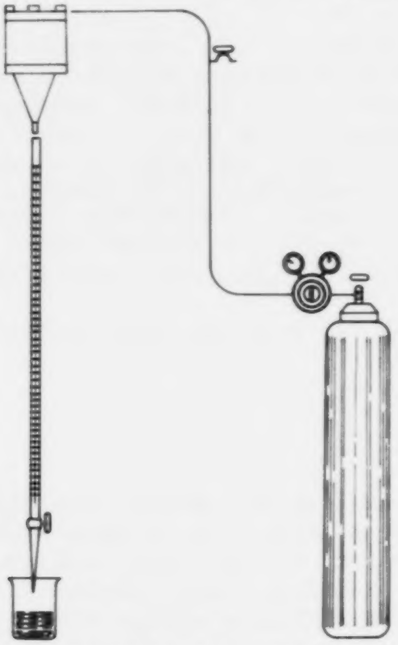

Figure C-2. Passavant Series 275 Resistance Meter.

Facilities

A Passavant Series 275 Resistance Meter was used. It included a 7 cm diameter stainless steel body and support screen. A 7 cm diameter filter cloth together with a standardized laboratory filter paper served as the filter media. Required additional equipment included a cylinder of compressed nitrogen gas, a 100 ml buret, stand, and stopwatch. The instrument was assembled as shown in Figure C-2.

Operation

The filter cloth was wetted and placed in the meter over the support screen. One of the laboratory filter papers was wetted and placed on top of this cloth. A 250 ml sample of the conditioned sludge was measured into the meter, and the top was attached and secured. The 100 ml buret was initially filled to its lower 100 ml mark; and the meter was then pressurized to 225 psig with the compressed gas. Simultaneous readings of filtrate quantity and time were subsequently recorded. Readings were taken at 15 second intervals for a period of 240 seconds or until the filtration was completed and blow-by occurred, with gas passing through the filter.

Analysis

The specific resistance equation was revised by Passavant Corp. to give:

$$R = Rp = \frac{2PA^2b}{\mu c} = \frac{Kb}{c}$$

In this equation, Rp is measured as a dimensionless quantity. K is an index constant developed by Passavant, measured in arbitrary units of g-cm^3/s. It is a function of the pressure, temperature, and viscosity of the sludge being dewatered. The slope, b (s/cm^6), of the filtrate curve and the weight of dry cake solids per unit volume of filtrate, c (g/cm^3), were obtained as described in the previous section. It was noted earlier that this pressure test is essentially a Buchner funnel test run under positive pressure. Results, therefore, could also be interpreted by the same formula used for the Buchner funnel test.

A sample of the data for this test and its analysis is shown in the accompanying data sheet.

CAPILLARY SUCTION METHOD

The use of this method for determining the filterability of sewage sludges was developed as the result of research studies at the Water Pollution Research Laboratory in Stevenage, England.[10,11] It was developed as an alternative to the modified Buchner funnel test to permit rapid assessment of the filterability of a sewage sludge.

The principle of the method is that filtration occurs by the suction applied to the sludge by the capillary action of a standard grade, absorbent filter paper. The rate at which the paper becomes wetted gives an indication of the filterability of the sludge.

Facilities

A capillary suction time meter, which consisted of two transparent plates separated by a filter paper and an automatic timer, was used. The lower plate measured 9 cm x 9 cm x 0.6 cm high. One edge of this plate was raised to a height of 1.2 cm and served to position the filter paper. (The type of filter paper used was the Whatman No. 17 chromatography grade). The upper plate measured 7 cm x 9 cm x 2.3 cm high and contained a central hole approximately 1.9 cm in diameter. On the under side of the plate, concentric with the center hole, were two circular marks of diameters 3.2 cm and 4.5 cm; these marks were connected electrically to an automatic timer. A stainless steel cylinder, 2.5 cm high and 1.8 cm inner diameter, fitted into the central hole of the upper plate and served as a reservoir for the sludge sample.

Operation

The filter paper was positioned on the lower plate along the raised edge. The upper plate, with electrical connections touching the filter paper, was placed on top of the filter paper along the raised edge of the lower plate. The stainless steel cylinder was placed in the hole in the upper plate and a small volume of the sludge sample poured into it. As the suction pressure of the filter paper drained the filtrate from the sample, the automatic timer started when the outward progression of the filtrate reached the first connection and stopped when it reached the second. The capillary suction time or CST (in seconds) was then read from the timer.

A picture of the CST instrument is shown in Figure C-3 and sample data is shown in the data sheet.

Analysis

The CST only provides an indication of the filterability of the sludge. Through calibration with the modified Buchner funnel and pressure methods, however, it can be correlated with the specific resistance parameter. (See Section 7, Special Tests).

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10. Gale, R.S. and R.C. Baskerville. "Capillary Suction Method for Determination of the Filtration Properties of a Solid/Liquid Suspension." Chemistry and Industry, 1967, p 355.

11. Gale, R.S. and R.C. Baskerville. "A Simple Automatic Instrument for determining the Filtrability of Sewage Sludges." Water Pollution Control, 1968, 67, p. 233.

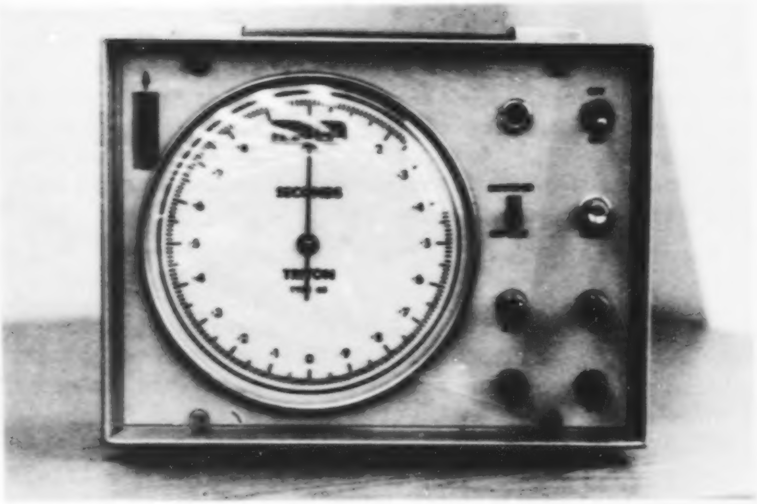


Figure C-3. CST Instrument

Date: 7/28/77 Run: #1

Sludge Description: Ratio Secondary/Primary @ 2/1

Conditioning Additives: $FeCl_3$ - 3.67 gals
Lime -11 gals

Test #1 CST = 10.9 sec.

Test #2 Pressure Method

Temperature: 29 °C Pressure: 225 psig

Time (θ) s	Reading (v) cm^3	Volume (V) cm^3	θ/V s/cm^3
0	100.0	0	-
15	84.0	16.0	0.938
30	75.8	24.2	1.240
45	68.8	31.2	1.442
60	63.2	36.8	1.630
75	58.0	42.0	1.786
90	53.2	46.8	1.923
105	48.6	51.4	2.043
120	45.0	55.0	2.182
135	41.0	59.0	2.288
150	37.2	62.8	2.389
165	34.0	66.0	2.500
180	31.0	69.0	2.609
195	27.8	72.2	2.701
210	24.6	75.4	2.785
225	21.8	78.2	2.877
240	18.8	81.2	2.956

Test # 3 Modified Buchner Funnel Method

Temperature: 29 °C Pressure: 24 " Hg

Time (θ) s	Volume (V) cm^3	θ/V s/cm^3
0	0	-
10	34	0.294
20	48	0.417
30	54	0.556
40	64	0.625
50	72	0.694
60	76	0.789
70	86	0.814
80	90	0.888
90	96	0.938
100	100	1.000
110	104	1.058
120	108	1.111
130	112	1.161
140	116	1.207
150	120	1.250
160	Vacuum broke @ 2' 38"	

ANALYSIS

Test # 1 CST = 10.9 sec

Test # 2 Pressure Method

a) Calculation of c

Unconditioned sludge:

$$21.0 \text{ inches} \times 6.4 \frac{\text{gal}}{\text{inch}} \times 3.785 \frac{\text{l}}{\text{gal}} = 500.76 \text{ l}$$

$$500.76 \text{ l} \times \frac{5.96\%}{100\%} \times 0.995 \frac{\text{g}}{\text{ml}} \times 1000 \frac{\text{ml}}{\text{l}} = 29725.65\text{g}$$

Lime:

$$11 \text{ gals} \times 3.785 \frac{\text{l}}{\text{gal}} = 41.64 \text{ l}$$

$$11 \text{ gals} \times 1 \frac{\text{lb}}{\text{gal}} \times 454 \frac{\text{g}}{\text{lb}} = 4994.0\text{g}$$

$FeCl_3$:

$$3.57 \text{ gals} \times 3.785 \frac{\text{l}}{\text{gal}} = 13.89 \text{ l}$$

$$3.67 \text{ gals} \times 1 \frac{\text{lb}}{\text{gal}} \times 454 \frac{\text{g}}{\text{lb}} = 1666.18\text{g}$$

	g	l
Sludge	29725.65	500.76
Lime	4994.00	41.64
$FeCl_3$	1666.18	13.89
	36385.83	556.29

$$\text{TSS} = \frac{36385.83}{556.29} = 65.41 \text{ g/l}$$

$$c = \frac{\text{TSS}}{1000 - \text{TSS}} = 0.070 \text{ g/cm}^3$$

b) @ 29 °C and 225 psig, $K = 5.43 \frac{\text{g-cm}^3}{\text{s}}$ (see chart)

c) from graph, $b = 0.0235 \text{ s/cm}^6$

d) $Rp = \frac{kb}{c}$

$$= \frac{5.43\ \text{g-cm}^3/\text{s} \times 0.0235\ \text{s/cm}^6}{0.070\ \text{g/cm}^3}$$

= 1.82 (dimensionless)

e. Alternatively, if Rp is analyzed according to the equation

$$Rp = \frac{2PA^2b}{\mu c}$$

$A = 38.5\ \text{cm}^2$; $A^2 = 1482.3\ \text{cm}^4$

$P = 225\ \text{psig} = 155.133 \times 10^4\ \text{N/m}^2$

$@29\ ^\circ\text{C}, \mu\ H_2O = 8.21 \times 10^{-4}\ \text{Ns/m}^2$

$$Rp = \frac{2PA^2b}{\mu c}$$

$$= \frac{2 \times 155.133 \times 10^4\ \text{N/m}^2 \times 1482.3\ \text{cm}^4 \times 0.0235\ \text{s/cm}^6}{8.21 \times 10^{-4}\ \text{Ns/m}^2 \times 0.070\ \text{g/cm}^3}$$

$= 1.88 \times 10^{12}\ \text{cm/g}$

Test #3 Modified Buchner Funnel Method

a) $c = 0.070\ \text{g/cm}^3$

b) $P = 24\text{"Hg} = 81360\ \text{N/m}^2$

c) $@\ 29^\circ\text{C}, \mu\ H_2O = 8.21\text{x}10^{-4}\ \text{Ns/m}^2$

d) $A = 63.6\ \text{cm}$; $A^2 = 4047.5\ \text{cm}^4$

e) from graph, $b = 0.00775\ \text{s/cm}^6$

f) $$Rv = \frac{2PA^2b}{\mu c}$$

$$= \frac{2 \times 81360\text{N/m}^2 \times 4047.5\ \text{cm}^4 \times 0.00775\text{s/cm}^6}{8.21\text{x}10^{-4}\ \text{Ns/m}^2 \text{x}\ 0.070\ \text{g/cm}^3}$$

$= 8.88 \times 10^{10}\ \text{cm/g}$

DATE- 7-28-77 PRESS- NGK SQUEEZING PRESSURE- 213 CLOTH TYPE NY 51-4

SLUDGE TYPE- Secondary/Primary APPEARANCE- Black and grainy

TEMPERATURE- GRINDING- Primary mazorated PURPOSE OF RUN- Comparison Study

RUN #	#1	#2	#3	#4	#5
PRIMARY/SECONDARY (pH)$_{pH}$	- / -	/	/	/	/
COMBINED SLUDGE-uncond.-cond.	- / -	/	/	/	/
LEVEL-before& after cond.(INCH)	27/24 1/2 /31 7/8	/	/	/	/
LEVEL (after pumping)	44 7/8				
LIME ADDED (GALS)	11				
FECL3 ADDED(GALS)	3.67				
SLUDGE PUMPED (GALS)					
FILTRATE COLLECTED (PUMPING)					
FILTRATE COLLECTED (SQUEEZE)					
FILTRATE COLLECTED (TOTAL)					
FILTRATE:pH					
FILTRATE APPEARANCE	yellowish				
CAKE:WEIGHT (WET)	105 #				
CAKE:CONSISTENCY/DISCHARGE	Excellent				
PUMPING TIME	17				
PUMPING PRESSURE (TERMINAL)	94				
SQUEEZING TIME	19				
CLOTH WASH: before run	yes				
CAKE THICKNESS	1/2 to 3/4				
CST (of conditioned sludge)	12.6				
CAKE DENSITY (g/cm^3)	1.175				
TANKED DRAINED	yes				
PRIMARY- SPECIFIC GRAVITY-	1.0154				
SECONDARY-SPECIFIC GRAVITY-	1.0				
COMBINED -UNCOND.(SP.GR.)	0.996				
COMBINED-COND. (SP.GR.)	1.0205				

PRIMARY-INCHES- 4 SECONDARY-INCHES- 17 RATIO-#/# PRI./SEC. 32/68

REMARKS: **Primary Solids = 10.8%**

Secondary Solids = 5.76%

Combined Solids = 5.96%

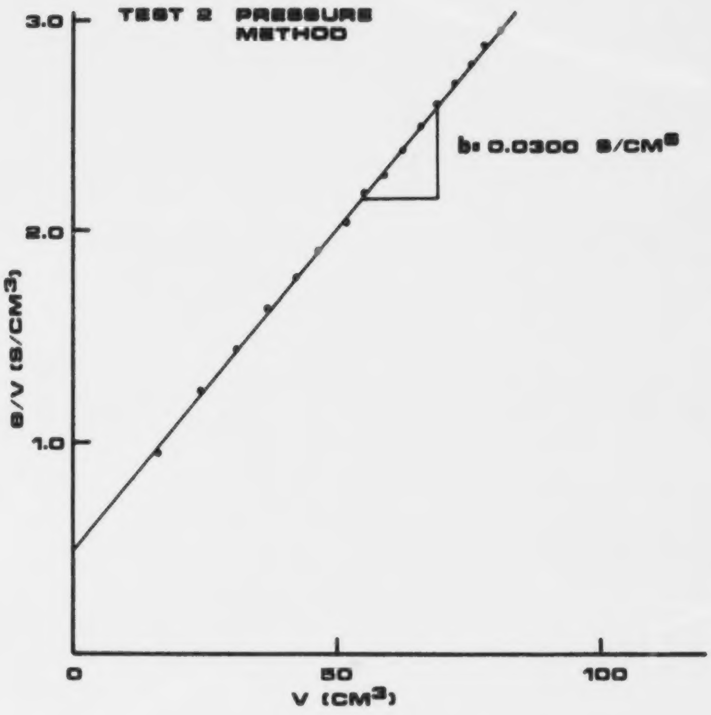
TABLE 7. EXTENDED RUNS - 3/8/77

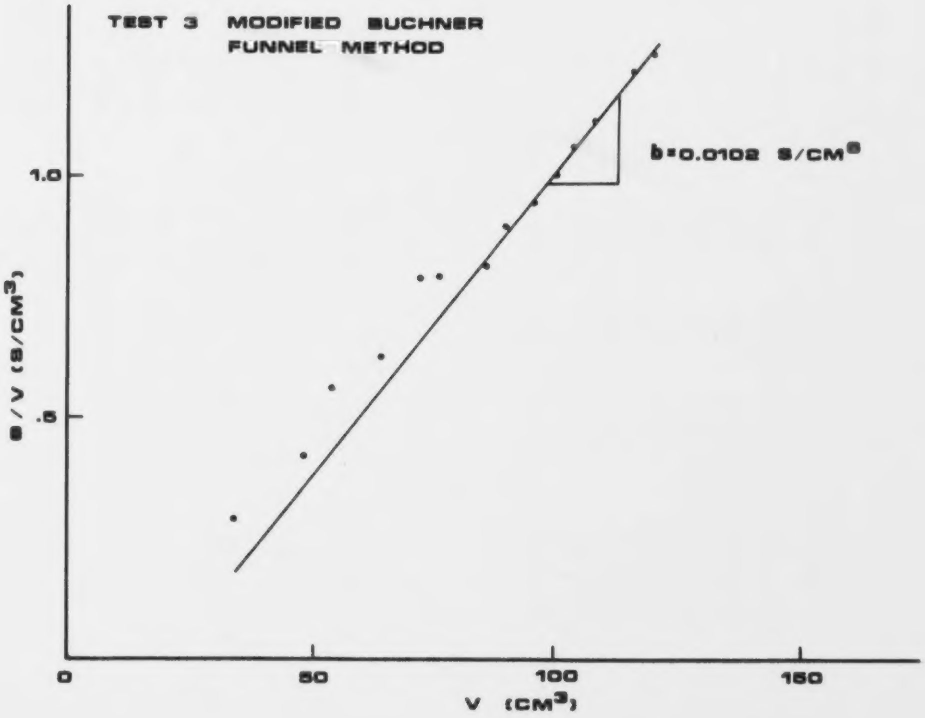
% CHEMICALS LIME/FeCl$_3$	pH	FILTRATE mg/l BOD$_5$	COD	PO$_4$	TKN	NH$_3$	NO$_3$	TOTAL ALKALINITY	TOTAL SOLIDS	SUSPENDED SOLIDS	
26.7/7.8	11.9	554	1560	35.8	173	85.3	.70	1801	7286	144	
26.8/7.8	11.9	916	2067	110	253	107	.74	2367	9431	1540	
26.8/7.8	11.9	694	2061	63.6	214	105	.80	1987	8575	464	
29.3/8.5	11.7	684	1640	44.7	244	106	.75	2021	8307	151	
29.3/8.5	11.7	682	1668	56.7	240	96.9	.77	1961	8203	62	
29.3/6.5	11.8	684	1785	69.2	296	99.2	.79	1959	8140	55	
30.0/8.8	11.8	420	1730	47.5	226	96.9	.74	2063	8471	127	
30.0/8.8	11.9	374	1867	64.0	253	110	.72	2008	8474	47	
28.5/8.3	11.8	626	1923	61.4	238	101	.75	2020	8361	323	(Average)

TABLE 8. FILTRATE QUALITY VS. CHEMICAL CONDITIONING

DATE	% CHEMICALS LIME/FeCl$_3$	% CAKE SOLIDS	CAKE DISCHARGE	TOTAL SOLIDS mg/l	SUSPENDED SOLIDS mg/l	pH
3-10-77	45.5/13.3	39.6	excellent	10120	87	11.5
3-10-77	34.4/10.0	38.5	excellent	8910	70	11.6
3-10-77	22.8/6.7	36.0	excellent	8189	80	11.5
3-10-77	17.1/5.0	31.6	good	9404	2604	11.5
7-12-77	16.8/5.6	41.8	excellent	9815/10130*	120/18*	11.5
7-12-77	14.4/4.9	39.9	excellent	9736/10024*	216/45*	11.5
7-12-77	12.2/4.1	29.0	poor	8963/8057*	438/198*	11.5
11-1-77	25.2/8.3	38.5	excellent	6773	28	-
11-1-77	19.7/6.6	36.8	excellent	5879	22	-
11-1-77	14.8/5.0	34.5	good	5546	189	-

* First entry is from the pump cycle; second entry from the squeezing cycle.


TEST 2 PRESSURE METHOD
b= 0.0300 S/CM^6
S/V (S/CM^3)
3.0
2.0
1.0
0
50
100
V (CM^3)


TEST 3 MODIFIED BUCHNER FUNNEL METHOD
b=0.0102 S/CM^6
S / V (S/CM^3)
1.0
.5
0
50
100
150
V (CM^3)

K-Factor as a Function of Temperature

Temperature		K	Temperature		K	Temperature		K
°C	°F		°C	°F		°C	°F	
0	32.0	2.48						
1	33.8	2.57	36	96.8	6.27	71	159.8	11.1
2	35.6	2.66	37	98.6	6.40	72	161.6	11.2
3	37.4	2.74	38	100.4	6.52	73	163.4	11.4
4	39.2	2.83	39	102.2	6.65	74	165.2	11.5
5	41.0	2.93	40	104.0	6.77	75	167.0	11.7
6	42.8	3.02	41	105.8	6.90	76	168.8	11.9
7	44.6	3.11	42	107.6	7.03	77	170.6	12.0
8	46.4	3.21	43	109.4	7.16	78	172.4	12.2
9	48.2	3.30	44	111.2	7.29	79	174.2	12.3
10	50.0	3.40	45	113.0	7.42	80	176.0	12.5
11	51.8	3.49	46	114.8	7.55	81	177.8	12.6
12	53.6	3.59	47	116.6	7.68	82	179.6	12.8
13	55.4	3.69	48	118.4	7.82	83	181.4	12.9
14	57.2	3.80	49	120.2	7.95	84	183.2	13.1
15	59.0	3.90	50	122.0	8.09	85	185.0	13.2
16	60.8	4.00	51	123.8	8.23	86	186.8	13.4
17	62.6	4.10	52	125.6	8.36	87	188.6	13.5
18	64.4	4.21	53	127.4	8.50	88	190.4	13.7
19	66.2	4.31	54	129.2	8.63	89	192.2	13.8
20	68.0	4.43	55	131.0	8.77	90	194.0	14.0
21	69.8	4.53	56	132.8	8.91	91	195.8	14.2
22	71.6	4.64	57	134.6	9.05	92	197.6	14.3
23	73.4	4.75	58	136.4	9.20	93	199.4	14.5
24	75.2	4.86	59	138.2	9.34	94	201.2	14.6
25	77.0	4.97	60	140.0	9.48	95	203.0	14.8
26	78.8	5.09	61	141.8	9.62	96	204.8	15.0
27	80.6	5.20	62	143.6	9.77	97	206.6	15.2
28	82.4	5.31	63	145.4	9.91	98	208.4	15.3
29	84.2	5.43	64	147.2	10.1	99	210.2	15.5
30	86.0	5.55	65	149.0	10.2	100	212.0	15.7
31	87.8	5.67	66	150.8	10.3			
32	89.6	5.79	67	152.6	10.5			
33	91.4	5.91	68	154.4	10.6			
34	93.2	6.03	69	156.2	10.8			
35	95.0	6.15	70	158.0	10.9			

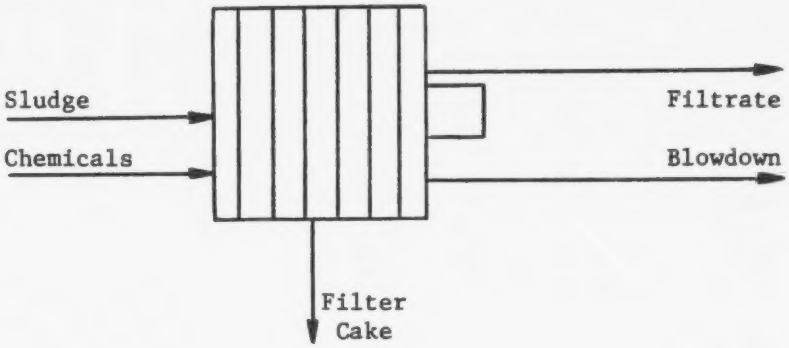
This table subsumes a constant pressure of 225 psig, a filtration area equal to that of the PASSAVANT Series 275 Resistance Meter, and the dynamic viscosity of the filtrate to be equivalent to that of water.

Source: Passavant Corporation

APPENDIX D

MATERIAL BALANCE

Data for this test was collected on 10/18/77. Refer to Data Sheet at end of this section.



INPUT, lbs total solids	
Sludge	37.5
Chemicals	10.0
Total	47.5

OUTPUT, lbs total solids	
Filter Cake	
Sludge	32.7
Chemicals	8.7
Filtrate	4.2
Sludge Blowdown	
Sludge	1.6
Chemicals	0.4
Filtrate Blowdown	0.2
Total	47.8

$$\text{Error in Analysis} = \frac{47.8-47.5}{47.5} \times 100 = 0.63\%$$

Calculations

1. Input

Solid input to the press consists of the sludge and chemical solids in the feed. The following parameters were measured:

i. feed volume to the press - 77.6 gals
ii. solid content of feed - 7.10%
iii. specific gravity of feed - 1.033
iv. chemical content of feed - 26.7% of dry sludge solids

The total mass of solids in the feed is calculated as:
Mass total solids = Mass Feed x % solids

$$= (77.6 \text{ gals} \times 8.345 \frac{\text{lbs}}{\text{gal}} \times 1.033) \times \frac{7.10\%}{100\%}$$

$$= 47.5 \text{ lbs}$$

and consists of both chemical and sludge solids.

The mass of chemical solids is:
Mass chemical solids = Mass total solids x % chemical solids

Chemical content of the feed is 26.7% of the dry sludge solids; therefore based on the total feed solids, i.e., sludge + chemicals, the chemical content is:

$$\frac{26.7}{126.7} \times 100 = 21.1\%$$

$$\text{Mass chemical solids} = 47.5 \text{ lbs} \times \frac{21.1\%}{100\%}$$

$$= 10.0 \text{ lbs}$$

The mass of dry sludge solids is:
Mass sludge solids = Mass total solids - Mass chemical solids
= 37.5 lbs

INPUT SUMMARY

chemical solids	10.0 lbs
sludge solids	37.5 lbs
	47.5 lbs

2. Output

Output from the press consists of solids in the cake, filtrate and blowdown. The following output variables were measured:

i. Filter cake weight - 101.0 lbs
ii. Solid content of filter cake - 41.0%
iii. Volume of filtrate collected - 58.3 gals
iv. Total solids concentration of filtrate - 8569 mg/l
v. Volume of blowdown - 6.4 gals

a. Filter Cake

Total mass of the cake is 101.0 lbs; total mass of the solids fraction of the cake is calculated as:

$$\text{Mass total solids} = \text{Mass cake} \times \%\ \text{solids}$$

$$= 101.0\ \text{lbs} \times \frac{41.0\%}{100\%}$$

$$= 41.4\ \text{lbs}$$

The mass of chemical solids in the cake is calculated as:
Mass chemical solids = Mass total solids x % chemical solids

$$= 41.4\ \text{lbs} \times \frac{21.1\%}{100\%}$$

$$= 8.7\ \text{lbs}$$

and the mass of sludge solids is:
Mass sludge solids = Mass total solids - Mass chemical solids

$$= 32.7\ \text{lbs}$$

b. Filtrate

A total volume of 58.3 gallons of filtrate was collected during the filter press cycle. The mass of solid particles within the filtrate is calculated as:
Mass filtrate solids = Volume filtrate x Concentration filtrate solids

$$= 58.3\ \text{gal} \times 8569\ \text{mg/l} \times 8.345 \times 10^{-6}\ \frac{\text{lb/gal}}{\text{mg/l}}$$

$$= 4.2\ \text{lbs}$$

c. Blowdown

The blowdown volume consists of sludge and filtrate collected during the blowdown cycle. It is assumed that the blowdown is composed of equal volumes of each.

Sludge Blowdown

The volume of sludge collected during blowdown is assumed to be 3.2 gallons. This sludge has a total mass of:

Mass sludge blowdown = Volume sludge blowdown x density sludge
= 3.2 gal x 8.345 lb/gal x 1.033
= 27.6 lbs

The total mass of solids within the blowdown is, therefore:

Mass total solids = Mass sludge blowdown x % solids

$$= 27.6 \text{ lbs} \times \frac{7.10\%}{100\%}$$

= 2.0 lbs

The mass of chemical solids is calculated as:

Mass chemical solids = Mass total solids x % chemical solids

$$= 2.0 \text{ lbs} \times \frac{21.1\%}{100\%}$$

= 0.4 lbs

and, the mass of sludge solids is:

Mass sludge solids = Mass total solids - Mass chemical solids
= 1.6 lbs

Filtrate Blowdown

The volume of filtrate collected is assumed to be 3.2 gallons. The mass of solid particles within the filtrate is, therefore, calculated as:

Mass filtrate solids = Volume filtrate x Concentration filtrate solids

$$= 3.2 \text{ gal} \times 8569 \text{ mg/l} \times 8.345 \times 10^{-6} \frac{\text{lb/gal}}{\text{mg/l}}$$

= 0.2 lbs

OUTPUT SUMMARY

Filter Cake	lbs
Sludge solids	32.7
Chemical solids	8.7
Filtrate	4.2
Sludge Blowdown	
Sludge solids	1.6
Chemical solids	0.4
Filtrate blowdown	0.2
Total	47.8

FILTER PRESS DATA

DATE- RUN# - 10/18/77				
TYPE SLUDGE-	Pri/Sec @ 1/2			
RATIO- #/#-PR./SEC.	30.7/69.3			
FEED SOLIDS-%SOL./%VOL$_T$(uncond.)	5.77/65.2			
PRIMARY-%SOL./%VOL/ pH.-	11.1 - 68.8			
SECONDARY-%SOL./%VOL./ pH.-	5.1 - 61.6			
FEED SOLIDS-%SOL./%VOL.-(cond.)	7.1 - 42.4			
PRIMARY- SP. GR. (gr./cc.)	1.034			
SECONDARY-SP.GR.-(gr./cc.)	1.014			
FEED SOLIDS-SP.GR.-(uncond.)- pH.-	1.008			
FEED SOLIDS-SP.GR.-(cond.)- pH.-	1.033			
LIME (added) %-	20.1			
FECL3 (added) %-	6.6			
VOLUME-(feed to press)-GALS.	77.6			
PUMP TIME-(Mins.)	14			
SQUEEZE TIME-(Mins.)-	15			
TERMINAL PRESSURE-Pump psig.-	75			
SQUEEZING PRESSURE-psig	213			
FILTRATE VOLUME-(gals.)PUMPING-	52.0			
FILTRATE VOLUME-(Gals.) SQUEEZE-	6.3 - 58.3			
FILTRATE VOL.-(Gals.) TOTAL+B.D.	6.4 - 64.7			
FILTRATE pH.	-			
FILTRATE (mg/1) TOTAL SOLIDS-	8569			
FILTRATE (mg/1) SUS. SOLIDS-	59			
FILTER CAKE-(Wet weight) lbs.	101.0			
FILTER CAKE-(%Sol./%Vol.)	41.0 - 48.3			
FILTER CAKE (Dry Weight) lbs.	32.7			
CAKE THICKNESS-(Inches)	3/8 - 3/4			
YIELD (lbs./ft.2hr.)	1.08			

CLOTH TYPE- NY 51-4 TEMPERATURE- - PRESS MECHANICAL TIME- 0 min. PRESS FILTER AREA- 62.4 ft^2

CONCLUSIONS-

APPENDIX E

DERIVATION OF COSTS

Costs are derived for a plant generating 250 dry tons/day (500,000 lbs/day) of sludge solids (roughly equivalent to a wastewater flow of 200-250 MGD).

VACUUM FILTER

Number of Units

Full-scale yield = 3.0 $\mathrm{lb/hr/ft^2}$
Filtration area = 600 $\mathrm{ft^2/unit}$

$$\text{Number of units} = \frac{500{,}000\ \mathrm{lb/day}}{3.0\ \mathrm{lb/hr/ft^2} \times 24\ \mathrm{hr/day} \times 600\ \mathrm{ft^2/unit}}$$

= 11.6 units
or 12 units + 1 spare = 13 units

Capital Costs

1. Filters--from Komline-Sanderson, the cost per unit is $158,000.

   Total installed cost = $158,000 x 13 units x 3 = $6,162,000

2. Lime System--for feeding an average of 50 tons/day

   Total installed cost = $1,000,000

   and includes conveyors, bins, slakers, pumps, etc.

3. Ferric Chloride System--for feeding 16.7 tons/day of 10% solution

   Total installed cost = $500,000

   and includes storage tanks, pumps, etc.

4. Conveyors--to transport filter cake to next process.

   Total cost = $1,000,000

5. Total Capital Costs--

Filters		$6,162,000
Lime		1,000,000
Ferric Chloride		500,000
Conveyors		1,000,000
	Total	$8,662,000

Annual Costs

1. Amortization--

Amortized cost = $8,700,000 x 0.09 = $783,000

2. Chemicals--chemical usage consists of lime @ 20%, ferric chloride @ 7%, and anti-scalant (to counteract lime scale).

Lime cost = 100,000 lb/day x $.022/lb x 365 days/yr = $803,000

$FeCl_3$ cost = 35,000 lb/day x $.065/lb x 365 days/yr = $830,375

Anti-scalant cost = $30,000

Total chemical costs = $1,663,375

3. Power--power costs assume 100% duty cycle usage.

Filters - 12 units @ 90 Hp/unit		1080 Hp
Sludge pumps - 12 @ 10 Hp/pump		120
$FeCl_3$ system		22
Lime system		87
Conveyors		37
	Total	1346 Hp

Power cost = 1346 Hp x .746 Kw/Hp x $.04/Kw-hr x 8760 hr/yr
= $ 351,900

4. Water--the cloth washing system requires 60 gpm/unit.

Water cost = 60 gpm/unit x 12 units x $.25/1000 gal x 525,600 min/yr
= $94,600

5. Operating Labor--labor costs assume each crew consists of 1 supervisor, 5 men to operate the filters and 1 man to operate the chemical system. To cover a 7 day/week operation, 4 crews will be required, and 28 man-years (7 men/crew x 4 crews) will be expended.

Labor cost = 28 man-years x $21,000/man-year = $588,000

6. Maintenance--maintenance costs consist of the costs for both normal maintenance (materials and labor) and filter cloth replacement. Normal maintenance is based on 2% of the purchase price of all equipment, i.e. (Total capital cost)/3. The filter cloths must be replaced once every 2000 hrs at a cost of $550 per cloth. The labor costs for changing the cloths are included in the cost for operating labor.

$$\text{Normal maintenance cost} = \frac{\$8{,}662{,}000}{3} \times \$.02 = \$58{,}000$$

$$\text{Cloth replacement cost} = \frac{8760 \text{ hr/yr}}{2000 \text{ hr/cloth/unit}} \times 12 \text{ units} \times \$550\text{/cloth}$$

$$= \$29{,}150$$

Total maintenance costs = $87,150

7. Total Annual Costs--

Amortization		$ 783,000
Chemicals		1,663,000
Power		351,900
Water		94,600
Labor		588,000
Maintenance		87,150
	Total	$3,567,650

Unit Cost

For processing 250 tons/day of dry sludge solids.

$$\text{Unit cost} = \frac{\$\ 3{,}567{,}650\text{/yr}}{250 \text{ tons/day} \times 365 \text{ days/yr}} = \$39.10\text{/ton}$$

FILTER PRESS--VARIABLE VOLUME UNIT

Number of Units

Full-scale yield = 0.49 lb/hr/ft^2
Filtration area = 5380 $\text{ft}^2\text{/unit}$

$$\text{Number of units} = \frac{500{,}000 \text{ lb/day}}{0.49 \text{ lb/hr/ft}^2 \times 24 \text{ hr/day} \times 5380 \text{ ft}^2\text{/unit}}$$

= 7.9 units
or 8 units + 1 spare = 9 units

Capital Costs

1. Presses--from Envirex, the cost for 9 units is \$6,500,000.

    Total installed cost = \$6,500,000 x 3 = \$19,500,000

2. Chemical System--for lime and $FeCl_3$; see vacuum filter costs.

    Total installed cost = \$1,500,000

3. Flight Conveyors--

    Total installed cost = \$2,000,000

4. Total Capital Costs--

Presses		\$19,500,000
Chemical System		1,500,000
Flight Conveyors		2,000,000
	Total	\$23,000,000

Annual Costs

1. Amortization--

    Amortized cost = \$23,000,000 x .09 = \$2,070,000

2. Chemicals--for lime, $FeCl_3$, and anti-scalant; see vacuum filter costs.

    Chemical cost = \$1,663,400

3. Power--costs assume 100% duty cycle usage

Press plus accessories--

Power usage = 37 Kw-hr/ton

Associated systems--

Lime system	81 Hp
$FeCl_3$ system	14
Sludge pumping-4 pumps @ 15 Hp ea.	60
Conditioning system	20
Conveyors	260
	435 Hp

$$\text{Power usage} = \frac{435\ \text{Hp} \times .746\ \text{Kw/Hp} \times 24\ \text{hr/day}}{250\ \text{ton/day}} = 31.2\ \text{Kw-hr/ton}$$

Power Costs--total power usage is 68.2 Kw-hr/ton

Power cost = 68.2 Kw-hr/ton x 250 ton/day x 365 days/yr x $.04/kw-hr
= $248,900

4. Water--filter cloths will require washing once every 20 press cycles. With a 54-minute cycle per press, the filter cloths will be washed 1.35 times per day and each wash will consume 3000 gallons. City water will be used.

Water cost = 3000 gal/cycle x 1.35 cycles/day/unit x 365 days/yr
x 8 units x $.5267/1000 gal
= $6400

5. Operating Labor--costs assume each crew consists of 1 supervisor, 4 men to operate the presses, and 1 man to operate the chemical system. Four crews will be required for a 7 day/week operation and 24 man-years will be expended.

Labor cost = 24 man-years x $21,000/man-year = $504,000

6. Maintenance--maintenance costs consist of the costs for normal equipment maintenance and filter cloth and diaphragm replacement. Normal equipment maintenance is based on 2% of the purchase price of all equipment. Filter cloths will require replacement once every 3000 cycles at a material cost of $6500/unit. Diaphragms will require replacement once every 20,000 cycles at a material cost of $26,000/unit. Operating labor will change the filter cloths and diaphragms; 1.5 man-years will be expended.

Normal maintenance cost = $23,000,000/3 x .02 = $153,000

$$\text{Cloth replacement cost} = \frac{\text{27 cycles/day}}{\text{3000 cycles}} \times \text{365 days/yr} \times \text{\$6500/unit}$$

$$\times \text{ 9 units}$$

$$= \text{\$195,000}$$

$$\text{Diaphragm replacement cost} = \frac{\text{27 cycles/day}}{\text{20,000 cycles}} \times \text{365 days/yr}$$

$$\times \text{ \$26,000/unit} \times \text{9 units}$$

$$= \text{\$117,000}$$

Labor cost = 1.5 man-years x $21,000/man-year = $31,500

Total maintenance cost = $496,500

7. Total Annual Costs

Amortization		\$ 2,070,000
Chemicals		1,663,400
Power		248,900
Water		6,400
Operating Labor		504,000
Maintenance		496,500
	Total	\$ 4,989,200

Unit Cost

For processing 250 tons/day of dry sludge solids

$$\text{Unit cost} = \frac{\$\ 4,989,200/\text{yr}}{250\ \text{tons/day} \times 365\ \text{days/yr}} = \$54.68/\text{ton}$$

FILTER PRESS--HIGH-PRESSURE FIXED VOLUME UNIT

Number of Units

Full-scale yield = 0.31 lb/hr/ft^2
Filtration area = 11625 ft^2/unit
Chamber size = 40 mm (1.57 inches)

$$\text{Number of units} = \frac{500,000\ \text{lb/day}}{0.31\ \text{lb/hr/ft}^2 \times 24\ \text{hr/day} \times 11625\ \text{ft}^2/\text{unit}}$$
$$= 5.8\ \text{units}$$
or 6 units + 1 spare = 7 units

Capital Costs

From Passavant, the cost for 7 units, including all chemical systems, conveyors, etc., is \$8,450,000.

Total installed cost = \$8,450,000 x 3 = \$25,350,000

Annual Costs

1. Amortization--

Amortized cost = \$25,350,000 x .09 = \$2,281,500

2. Chemicals--for lime, $FeCl_3$, and anti-scalant; see vacuum filter costs.

Chemical cost = \$1,663,400

3. Power--for 6 presses in operation, power usage is 57.3 Kw-hr/ton

Power cost = 57.3 Kw-hr/ton x 250 tons/day x 365 days/yr x \$.04/Kw-hr
= \$210,000

4. Water--filter cloths will require washing once per month; each wash will consume 36,000 gallons of water.

Water cost = 36,000 gal/cycle x 1 cycle/mo/unit x 12 mo/yr

x 7 units x \$ .5267/1000 gal

= \$1600

5. Operating Labor--costs assume each crew consists of 1 supervisor, 3 men to operate the presses, and 1 man to operate the chemical system. For a 7 day/week operation, 20 man-years will be expended.

Labor cost = 20 man-years x \$21,000/man-year = \$420,000

6. Maintenance--maintenance costs consist of the costs for normal equipment maintenance and filter cloth replacement. Equipment maintenance is based on 2% of the purchase price of all equipment. Filter cloths will require replacement once per year at a cost of \$240,000 for materials and \$40,000 for labor (includes 2 men on day shift year-round).

Normal maintenance cost = \$25,350,000/3 x .02 = \$170,000

Cloth replacement cost = \$240,000

Labor cost = \$40,000

Total maintenance costs = \$450,000

7. Total Annual Costs--

Amortization	\$2,281,500
Chemicals	1,663,400
Power	210,000
Water	1,600
Operating Labor	420,000
Maintenance	450,000
Total	\$5,026,500

Unit Cost

For processing 250 tons/day of dry sludge solids.

$$\text{Unit cost} = \frac{\$5,026,500/\text{yr.}}{250 \text{ tons/day} \times 365 \text{ days/yr}} = \$55.08/\text{ton}$$

FILTER PRESS - LOW-PRESSURE FIXED VOLUME UNIT

Number of Units

Full-scale yield = 0.22 lb/hr/ft^2
Filtration area = 6760 ft^2/unit
Chamber size = 32 mm (1.25 inches)

$$\text{Number of units} = \frac{500{,}000 \text{ lb/day}}{0.22 \text{ lb/hr/ft}^2 \times 24 \text{ hr/day} \times 6760 \text{ ft}^2\text{/unit}}$$

= 14 units
or 14 units + 1 spare = 15 units

Capital Costs

1. Presses--from Nichols, the cost for each press is $400,000.

   Total installed cost = $400,000/unit x 15 units x 3 = $18,000,000

2. Chemical System--for lime and FeCl$_3$, see vacuum filter costs.

   Total installed cost = $1,500,000

3. Conveyors--

   Total installed cost = $2,300,000

4. Total Capital Costs--

Presses		$18,000,000
Chemical System		1,500,000
Conveyors		2,300,000
	Total	$21,800,000

Annual Costs

1. Amortization--

   Amortized cost = $21,800,000 x .09 = $1,962,000

2. Chemicals--for lime, FeCl$_3$, and anti-scalant; see vacuum filter costs.

   Chemical cost = $1,663,400

3. Power--

Press--includes sludge and chemical feed systems.

Power usage = 65.3 Kw-hr/ton

Transfer conveyors--power usage is 330 Hp.

$$\text{Power usage} = \frac{330 \text{ Hp} \times .746 \text{ Kw/Hp} \times 24 \text{ hr/day}}{250 \text{ tons/day}} = 23.6 \text{ Kw-hr/ton}$$

Power costs - total power usage is 88.9 kw-hr/ton

Power cost = 88.9 Kw-hr/ton x 250 tons/day x 365 days/yr x $.04/Kw-hr
= $324,500

4. Water--filter cloths on one press only will be washed each day. Each wash will consume 5000 gallons of water.

Water cost = 5000 gal/cycle x 1 cycle/day x 365 days/yr
x $.5267/1000 gal = $1,000

5. Operating Labor--costs assume each crew consists of 1 supervisor, 7 press operators, and 1 man to operate the chemical system. For a 7 day/week operation, 36 man-years will be expended.

Labor cost = 36 man-years x $21,000/man-year = $756,000

6. Maintenance--maintenance costs consist of the costs for normal equipment maintenance and cloth replacement. Equipment maintenance is based on 2% of the purchase price of all equipment. Filter cloths will require replacement once per year at a cost of $4600/unit for materials and $6,000 for labor (600 man-hr/yr).

Normal maintenance cost = $21,500,000/3 x .02 = $143,000

Cloth replacement cost = $4600/unit x 15 units = $69,000

Labor cost = $6,000

Total maintenance costs = $218,000

7. Total Annual Costs--

Amortization	$1,962,000
Chemicals	1,663,000
Power	324,500
Water	1,000
Operating Labor	756,000
Maintenance	218,000
Total	$4,924,500

Unit Cost

To process 250 tons/day of dry sludge solids.

$$\text{Unit cost} = \frac{\$4{,}924{,}500/\text{yr}}{250 \text{ tons/day} \times 365 \text{ days/yr}} = \$53.97/\text{ton}$$

BELT PRESS

Number of Units

Full-scale yield = 675 lb/hr/m width
Belt width = 3m/unit

$$\text{Number of units} = \frac{500{,}000 \text{ lb/day}}{675 \text{ lb/hr/m} \times 24 \text{ hr/day} \times 3\text{m/unit}} = 10.3 \text{ units}$$

or 11 + 1 spare=12 units.

Capital Costs

1. Presses--from Komline Sanderson, the cost per unit is $147,000.

   Total installed cost = $147,000/unit x 12 units x 3
   = $5,300,000

2. Polymer Feed System - includes storage, mixing, pumping, etc.

   Total installed cost = $750,000

3. Conveyors--

   Total installed cost = $1,000,000

4. Total Capital Costs--

Presses		$5,300,000
Chemical System		750,000
Conveyors		1,000,000
	Total	$7,050,000

Annual Costs

1. Amortization--

   Amortized cost = $7,050,000 x .09 = $634,500

2. Chemicals--test work indicated $9.00 per ton of sludge processed. However, because of uncertainty of polymer suitability, assume $15.00 per ton.

   Chemical cost = 250 tons/day x 365 days/yr x $15.00/ton
   = $1,368,800

3. Power--

Press - 12.75 Hp/unit x 11 units	140 Hp
Sludge pumps - 10 Hp/unit x 11 units	110
Polymer system	31
Conveyors	30
Total	311 Hp

Power cost = 311 Hp x .746 Kw-hr/Hp x 8760 hr/yr x \$.04/Kw-hr
= \$81,300

4. Water--each unit will consume 75 gpm.

Water cost = 75 gal/min/unit x 11 units x 525,600 min/yr
x \$.25/1000 gal
= \$108,400

5. Labor--costs assume each crew consists of 1 supervisor, 6 men to operate the presses, and 1 man to operate the chemical system. For a 7 day/week operation, 32 man-years will be expended.

Labor cost = 32 man-years x \$21,000/man-year = \$672,000

6. Maintenance--mantenance costs consist of the costs for normal maintenance and belt replacement. Normal maintenance costs for materials and labor are based on 3% of the purchase price of all equipment. Belt replacement costs will total \$20,000 per year.

Normal maintenance cost = \$7,050,000/3 x .03 = \$70,500

Belt replacement cost = \$20,000

Total maintenance costs = \$90,500

7. Total Annual Costs--

Amortization	\$ 634,500
Chemicals	1,368,800
Power	81,300
Water	108,400
Operating Labor	672,000
Maintenance	90,500
	\$2,955,500

Unit Cost

For processing 250 tons/day of dry sludge solids.

$$\text{Unit cost} = \frac{\$2,955,500}{250 \text{ tons/day} \times 365 \text{ days/yr}} = \$32.39/\text{ton}$$

VACUUM FILTER PLUS BELT PRESS

Number of Units

Vacuum filters - 12 units + 1 spare = 13 units
Full-scale yield for belt press = 1181 lb/hr/m width (Parkson tests)
Belt width - 2 m/unit

$$\begin{aligned}\text{Number of units} &= \frac{500{,}000\ \text{lb/day}}{1181\ \text{lb/hr/m} \times 24\ \text{hr/day} \times 2\text{m/unit}}\\ &= 8.8\ \text{units}\\ &\text{or } 9\ \text{units} + 1\ \text{spare} = 10\ \text{units}\end{aligned}$$

Capital Costs

1. Vacuum Filters--

Total installed cost = $8,700,000

2. Belt Presses--from Parkson, the cost per unit is $72,000

Total installed cost = $72,000/unit x 10 units x 3
= $2,160,000

3. Distribution and Feeding System--

Total installed cost = $1,000,000

4. Additional Conveyors--

Total installed cost = $500,000

5. Total Capital Costs--

Vacuum Filters	$8,700,000
Belt Presses	2,160,000
Distribution/Feed System	1,000,000
Additional Conveyor	500,000
Total	$12,400,000

Annual Costs

1. Amortization--

Amortized cost = $12,400,000 x .09 = $1,116,000

2. Chemicals--same as costs for vacuum filters.

Chemical cost = $1,663,400

3. Power--

Vacuum Filter System	1346 Hp
Belt Presses - 12.5 Hp/unit x 9 units	112.5
Distribution/Feed System	50
Additional Conveyors	15
Total	1523.5 Hp

Power cost = 1523.5 Hp x .746 Kw/Hp x $.04/Kw-hr
x 8760 hr/yr

= $398,240

4. Water--the vacuum filter system will consume 60 gpm/unit; the belt press system will consume 50 gpm/unit.

Water consumption = 60 gpm/unit x 12 units
+ 50 gpm/unit x 9 units
= 1170 gpm

Water cost = 1170 gal/min x 525,600 min/yr x $.25/1000 gal
= $153,700

5. Operating Labor - costs assume a 7-man crew will operate the vacuum filter system and a 3-man crew will operate the belt press system. For a 7 day/week operation, 40 man-years will be expended.

Labor cost = 40 man-years x $21,000/man-year
= $840,000

6. Maintenance--maintenance costs consist of normal maintenance costs on both the vacuum filters and the belt presses and belt replacement costs on the belt press. Vacuum filter maintenance costs will total $87,150 per year (see vacuum filter costs). Belt press maintenance costs are based on 3% of the purchase price of the additional equipment associated with the belt press. Belt replacement costs will total $20,000 per year.

Vacuum filter maintenance costs = $87,150

Belt press maintenance costs = $3,660,000/3 x .03
= $36,600

Belt replacement costs = $20,000

Total maintenance costs = $143,750

7. Total Annual Costs--

Amortization	$1,116,000
Chemicals	1,663,400
Power	398,240
Water	153,700
Operating Labor	840,000
Maintenance	143,750
	$4,315,090

Unit Cost

For processing 250 tons/day of sludge solids.

$$\text{Unit cost} = \frac{\$4,315,090}{250 \text{ tons/day} \times 365 \text{ days/yr}} = \$47.29/\text{ton}$$

INCINERATION

The costs in this section are rough approximations developed from on-going design work for the District of Columbia.

Number of Units

Incinerator rating = 10 lb wet feed/hr/ft^2 of burning area
Burning area for a 12 hearth unit = 4584 ft^2/unit (25.75 ft diameter)
Feed capacity = 45,480 lb wet feed/hr
Feed rate = 317.5 tons/day of dry solids (250 tons/day of dry sludge solids + 27% chemicals)
Availability factor = 85%

For a 20% feed

$$\text{Number of units} = \frac{317.5 \text{ tons/day} \times 2000 \text{ lbs/ton}}{\frac{45,480 \text{ lb wet feed}}{\text{hr/unit}} \times \frac{.2 \text{ lb dry feed}}{\text{lb wet feed}} \times \frac{24 \text{ hr}}{\text{day}} \times 0.85}$$

$$= 3.4 \text{ or } 4 \text{ units}$$

Similarly, for a 35% feed

Number of units = 1.9 or 2 units

Capital Costs

Includes air pollution control equipment (electrostatic precipitator) to meet emission requirements, installation, building, utilities, and engineering.

Total installed cost = $5,000,000/unit

Annual Costs

1. Amortization--

For a 20% feed, amortized cost = \$20,000,000 x .09 = \$1,800,000

For a 35% feed, amortized cost = \$900,000

2. Power--

For a 20% feed, power usage is 860 Hp/unit

Power cost = 860 hp/unit x 4 units x .746Kw/Hp x 8760 hr/yr
x \$ .04/Kw-hr
= \$899,200

For a 35% feed, power usage is 775 Hp/unit

Power cost = \$405,200

3. Fuel--the incinerator will produce an 800 °F outlet temperature. A fume furnace will raise all stack gases to 1350 °F before discharge. Water vapor will be removed in a subcooler, prior to reheating the stack gases. With a 20% solids feed, 21,000 gal/day of #2 fuel oil will be required for incineration, and 7,133 gal/day of #2 fuel oil will be required for the fume furnace, for a total fuel usage of 28,133 gal/day. With a 35% solids feed, 4306 gal/day of #2 fuel oil will be required for the fume furnace only.

For a 20% feed, fuel cost = 28,133 gal/day x 365 days/yr x \$ .40/gal
= \$4,110,000

For a 35% feed, fuel cost = \$630,000

4. Operating Labor--costs assume each crew consists of 1 supervisor, 1 operator per unit and 1 helper. Four crews will be required to cover a 7 day/week operation. For a 20% feed, 24 man-years will be expended; for a 35% feed, 16 man-years will be expended.

For a 20% feed, labor cost = 24 man-years x \$21,000/man-year
= \$504,000

For a 35% feed, labor cost = \$336,000

5. Maintenance--costs assume \$100,000/unit per year.

For a 20% feed, maintenance cost = \$100,000/unit x 4 units
= \$400,000

For a 35% feed, maintenance cost = \$200,000

6. Ash Disposal--hauling and disposal costs will total $10.00/ton of ash and are computed based on 40% of the incoming feed to the dewatering process plus 100% of the inert chemicals added.

Total ash quantity = 167.5 tons/day

Ash disposal costs = 167.5 ton/day x 365 days/yr x $10.00/ton
= $610,000

7. Total Annual Costs--

	20% Feed	35% Feed
Amortization	$1,800,000	$ 900,000
Power	899,200	405,200
Fuel	4,110,000	630,000
Operating Labor	504,000	336,000
Maintenance	400,000	200,000
Ash Disposal	610,000	610,000
Total	$8,323,200	$3,081,200

Unit Cost

For processing 250 tons/day of sludge solids.

For a 20% feed, unit cost = $\frac{\$8,323,200/\text{yr}}{250 \text{ tons/day} \times 365 \text{ days/yr}}$ = $91.21/ton

For a 35% feed, unit cost = $33.77/ton

HAULING

Hauling costs are based on actual costs now incurred at Blue Plains to haul sludge cake a 25 mile distance. Undigested vacuum-filter cake must be transported in enclosed vehicles, e.g., a concrete mixer; hence costs are $9.40 per wet ton. Filter-press cake is assumed to be dry enough to carry in an open dump truck; hence costs of $6.25 per wet ton.

The costs per dry ton of sludge solids were developed by correcting the above figures for the percent cake solids and quantities of chemicals added. For example, the cost of hauling vacuum filter cake at 20% solids is calculated as

$$\text{Hauling cost} = \frac{\$9.40/\text{wet ton}}{0.2 \text{ dry ton/wet ton}} \times \frac{1.27 \text{ tons total solids}}{1.0 \text{ ton sludge solids}}$$

$$= \$59.69/\text{dry ton of sludge solids}$$

COMPOSTING

Costs were obtained from a paper entitled "Composting Filter Press Cake"; presented at Compost Science Meeting, April, 1978 at Omaha, Nebraska; G. Wilson, D. Colacicco, and D. Casey, USDA, Beltsville, Maryland. Costs in this paper are presented in $/wet ton of sludge as received. To convert to $/dry ton of sludge solids, use the procedure as described under hauling costs.

APPENDIX F
FULL-SCALE UNIT SPECIFICATIONS
TABLE F-1. FILTER PRESS SPECIFICATIONS

	NGK	LASTA	PASSAVANT	NICHOLS
Filtration area, m^2	500	204	1080	628
No. of Chambers	130	32	150	115
Plate dimensions, m x m	1.5x1.5	2x2	2x2	2x1.5
Chamber thickness, mm	30	25	40	32
Yield, $lb/hr/ft^2$	0.49	0.60	0.31	0.22
Avg cycle time, min	54	∼40	210	180
No. of units for 250 TPD*	9	17	7	15
Budget purchase price, $1000/unit	722	775	715	400
Power usage, Kw-hr/ton	68.2	62.5	57.3	88.9

*Includes 1 spare

TABLE F-2. FILTER MEDIA SPECIFICATIONS

TYPE	CONSTRUCTION	MATERIAL	AIR PERMEABILITY at Δ P = 12.7 mm H_2O $cm^3/sec/cm^2$
NGK NY516	plain	polyamide/polypropylene	4.0
NGK TR520	herringbone twill	polyester/polyester	11.0
NGK NY51-4	twill	polyamide/polyester	93.0
LASTA P920	2x2 twill	polypropylene	13.3
LASTA P891	2x2 twill	polypropylene	25.0
LASTA P940	2x2 twill	polypropylene	40.0
PASSAVANT T167	twill	nylon	76.7
NICHOLS 4709/40	2x2 twill	polypropylene	20.3

GLOSSARY OF TERMS

Sludge Solids:	Sewage sludge solids only.
Total Solids:	Sewage sludge plus chemical solids.
Process Yield:	Calculated as kilograms of sludge solids per hour of filtration time per square meter of filtration area.
Full-scale Yield:	Same calculation as process yield except that cycle time includes both filtration and mechanical cycle time for a full-scale press.

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